U.S. DEPARTMENT OF COMMERCE
National Technical Information Service

PB-255 317

Proceedings: Symposium on Flue Gas Desulfurization - New Orleans, March 1976. Volume I

Industrial Environmental Research Lab.

May 1976

KEEP UP TO DATE

Between the time you ordered this report—which is only one of the hundreds of thousands in the NTIS information collection available to you—and the time you are reading this message, several *new* reports relevant to your interests probably have entered the collection.

Subscribe to the **Weekly Government Abstracts** series that will bring you summaries of new reports *as soon as they are received by NTIS* from the originators of the research. The WGA's are an NTIS weekly newsletter service covering the most recent research findings in 25 areas of industrial, technological, and sociological interest—invaluable information for executives and professionals who must keep up to date.

The executive and professional information service provided by NTIS in the **Weekly Government Abstracts** newsletters will give you thorough and comprehensive coverage of government-conducted or sponsored research activities. And you'll get this important information within two weeks of the time it's released by originating agencies.

WGA newsletters are computer produced and electronically photocomposed to slash the time gap between the release of a report and its availability. You can learn about technical innovations immediately—and use them in the most meaningful and productive ways possible for your organization. Please request NTIS-PR-205/PCW for more information.

The weekly newsletter series will keep you current. But *learn what you have missed in the past* by ordering a computer **NTISearch** of all the research reports in your area of interest, dating as far back as 1964, if you wish. Please request NTIS-PR-186/PCN for more information.

WRITE: Managing Editor
5285 Port Royal Road
Springfield, VA 22161

Keep Up To Date With SRIM

SRIM (Selected Research in Microfiche) provides you with regular, automatic distribution of the complete texts of NTIS research reports *only* in the subject areas you select. SRIM covers almost all Government research reports by subject area and/or the originating Federal or local government agency. You may subscribe by any category or subcategory of our WGA **(Weekly Government Abstracts)** or **Government Reports Announcements and Index** categories, or to the reports issued by a particular agency such as the Department of Defense, Federal Energy Administration, or Environmental Protection Agency. Other options that will give you greater selectivity are available on request.

The cost of SRIM service is only 45¢ domestic (60¢ foreign) for each complete microfiched report. Your SRIM service begins as soon as your order is received and processed and you will receive biweekly shipments thereafter. If you wish, your service will be backdated to furnish you microfiche of reports issued earlier.

Because of contractual arrangements with several Special Technology Groups, not all NTIS reports are distributed in the SRIM program. You will receive a notice in your microfiche shipments identifying the exceptionally priced reports not available through SRIM.

A deposit account with NTIS is required before this service can be initiated. If you have specific questions concerning this service, please call (703) 451-1558, or write NTIS, attention SRIM Product Manager.

This information product distributed by

NTIS U.S. DEPARTMENT OF COMMERCE
National Technical Information Service
5285 Port Royal Road
Springfield, Virginia 22161

PB 255 317

212150

EPA-600/2-76-138a
May 1976

Environmental Protection Technology Series

[illegible]

UNITED STATES ENVIRONMENTAL PROTECTION AGENCY

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U. S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

RESEARCH REPORTING SERIES

Research reports of the Office of Research and Development, U.S. Environmental Protection Agency, have been grouped into five series. These five broad categories were established to facilitate further development and application of environmental technology. Elimination of traditional grouping was consciously planned to foster technology transfer and a maximum interface in related fields. The five series are:

1. Environmental Health Effects Research
2. Environmental Protection Technology
3. Ecological Research
4. Environmental Monitoring
5. Socioeconomic Environmental Studies

This report has been assigned to the ENVIRONMENTAL PROTECTION TECHNOLOGY series. This series describes research performed to develop and demonstrate instrumentation, equipment, and methodology to repair or prevent environmental degradation from point and non-point sources of pollution. This work provides the new or improved technology required for the control and treatment of pollution sources to meet environmental quality standards.

EPA REVIEW NOTICE

This report has been reviewed by the U.S. Environmental Protection Agency, and approved for publication. Approval does not signify that the contents necessarily reflect the views and policy of the Agency, nor does mention of trade names or commercial products constitute endorsement or recommendation for use.

This document is available to the public through the National Technical Information Service, Springfield, Virginia 22161.

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/2-76-136a	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Proceedings: Symposium on Flue Gas Desulfurization--New Orleans, March 1976; Volume I		5. REPORT DATE May 1976 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) R.D. Stern, Chairman; W.H. Ponder and R.C. Christman (TRW, Inc.), Vice-chairmen		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS Miscellaneous		10. PROGRAM ELEMENT NO. EHE624 11. CONTRACT/GRANT NO. In-house
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Industrial Environmental Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Proceedings; 3/8-11/76 14. SPONSORING AGENCY CODE EPA-ORD

15. SUPPLEMENTARY NOTES IERL-RTP project officers for these proceedings are R.D. Stern and W.H. Ponder, Mail Drop 61, Ext 2915.

16. ABSTRACT The proceedings document the presentations made during the symposium, which dealt with the status of flue gas desulfurization technology in the United States and abroad. Subjects considered included: regenerable, non-regenerable, and advanced processes; process costs; and by-product disposal, utilization, and marketing. The purpose of the symposium was to provide developers, vendors, users and those concerned with regulatory guidelines with a current review of progress made in applying processes for the reduction of sulfur dioxide emissions at the full- and semi-commercial scale.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution	Byproducts	Air Pollution Control	13B
Flue Gases	Disposal	Stationary Sources	21B
Desulfurization	Utilization		07A,07D
Sulfur Dioxide	Marketing		07B
Sulfur Oxides			
Cost Effectiveness			14A

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 562
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE 13.50

EPA Form 2220-1 (9-73)

EPA-600/2-76-136a
May 1976

PROCEEDINGS:
SYMPOSIUM ON FLUE GAS DESULFURIZATION
NEW ORLEANS, MARCH 1976
VOLUME I

Program Element No. EHE624

Chairman: Richard D. Stern
Vice-Chairmen: Wade H. Ponder and Roger C. Christman

Industrial Environmental Research Laboratory
Office of Energy, Minerals, and Industry
Research Triangle Park, NC 27711

Prepared for

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, DC 20460

PREFACE

More than half of all "man-made" sulfur dioxide (SO_2) is emitted by electric power plants, and the use of sulfur-containing fossil fuels to generate electricity is predicted to increase by 50 percent by 1985. As a result, the development of SO_2 control technologies is one of the most important goals of the U.S. Environmental Protection Agency (EPA). Flue gas desulfurization (FGD) is the most promising technique for control of SO_2 that will be available for widespread application to fossil fuel-fired electric power plants for at least the next decade.

The Industrial Environmental Research Laboratory - Research Triangle Park (IERL-RTP) of EPA's Office of Research and Development sponsors symposia for the transfer of information regarding FGD research, development and application activities with the objective of further accelerating the development and commercialization of this technology. These symposia provide an opportunity for users and developers to discuss their experiences and the status of development and application of FGD technology.

The March 1976 symposium addressed full-scale FGD process applications in the U.S. and Japan as well as laboratory, pilot, and prototype research and development efforts. The symposium also provided an opportunity for the announcement of data and results which were previously unreported or not widely publicized. The economics of FGD and the disposal, utilization, and marketing of FGD system by-products were also discussed. The symposium papers were presented by a cross-section of those concerned with FGD including users, government and private developers, and vendors. The electric utility industry--the principal user of FGD--participated extensively in the symposium program. More than 650 people attended the symposium.

These Proceedings are comprised of copies of the participating authors' papers as received. As supplies permit, copies of the Proceedings are available free of charge and may be obtained by contacting IERL-RTP's Technical Information Coordinator, Environmental Protection Agency, Research Triangle Park, North Carolina 27711.

CONTENTS

TITLE PAGE

VOLUME I

OPENING SESSION 1

Keynote Address - SULFUR OXIDE CONTROL AND ELECTRICITY PRODUCTION
R. E. Train, Environmental Protection Agency, Washington, D. C. 3

REMARKS
S. J. Gage, Environmental Protection Agency, Washington, D. C. 9

STATUS OF FLUE GAS DESULFURIZATION SYSTEMS IN THE UNITED STATES
T. W. Devitt, G. A. Isaacs, and B. A. Laseke, PEDCo-Environmental Specialists, Inc., Cincinnati, Ohio 13

STATUS OF FLUE GAS DESULFURIZATION AND SIMULTANEOUS REMOVAL OF SO_2 AND NO_x IN JAPAN
J. Ando, Chuo University, Tokyo, Japan 53

FLUE GAS DESULFURIZATION ECONOMICS
G. G. McGlamery, H. L. Faucett, R. L. Torstrick, and L. J. Henson, Tennessee Valley Authority, Muscle Shoals, Alabama 79

STATUS OF THE EPRI FLUE GAS DESULFURIZATION DEVELOPMENT PROGRAM
L. W. Nannen and K. E. Yeager, Electric Power Research Institute, Palo Alto, California 101

NON-REGENERABLE PROCESSES SESSION 115

IERL-RTP SCRUBBER STUDIES RELATED TO FORCED OXIDATION
R. H. Borgwardt, Environmental Protection Agency, Research Triangle Park, North Carolina 117

RESULTS OF MIST ELIMINATION AND ALKALI UTILIZATION TESTING AT THE EPA ALKALI SCRUBBING TEST FACILITY
M. Epstein, H. N. Head, S. C. Wang, and D. A. Burbank, Bechtel Corporation, San Francisco, California 145

TITLE PAGE

DUQUESNE LIGHT COMPANY ELRAMA AND PHILLIPS POWER STATIONS LIME SCRUBBING FACILITIES
R. G. Knight and S. L. Pernick, Duquesne Light Company, Pittsburgh, Pennsylvania 205

OPERATIONAL STATUS AND PERFORMANCE OF THE COMMONWEALTH EDISON COMPANY WILL COUNTY LIMESTONE SCRUBBER
W. G. Stober, Commonwealth Edison Company, Chicago, Illinois 219

MHI FLUE GAS DESULFURIZATION SYSTEMS APPLIED TO SEVERAL EMISSION SOURCES
M. Hirai, M. Atsukawa, A. Tatani, and K. Kondo, Mitsubishi Heavy Industries, Ltd., Tokyo, Japan 249

STATUS OF FLUE GAS DESULFURIZATION USING ALKALINE FLY ASH FROM WESTERN COALS
H. M. Ness, E. A. Sondreal, and P. H. Tufte, U. S. Energy Research and Development Administration, Grand Forks, North Dakota 269

RESULTS OF THE 170 MW TEST MODULES PROGRAM, MOHAVE GENERATING STATION, SOUTHERN CALIFORNIA EDISON COMPANY
A. Weir, Jr., L. T. Papay, D. G. Jones, J. M. Johnson, and W. C. Martin, Southern California Edison Company, Rosemead, California 325

LA CYGNE STATION UNIT NO. 1 WET SCRUBBER OPERATING EXPERIENCE
C. F. McDaniel, Kansas City Power and Light Company, La Cygne, Kansas 355

RECENT SCRUBBER EXPERIENCE AT THE LAWRENCE ENERGY CENTER, THE KANSAS POWER AND LIGHT COMPANY
D. M. Miller, Kansas Power and Light Company, Topeka, Kansas 373

INTRODUCTION TO DOUBLE ALKALI FLUE GAS DESULFURIZATION TECHNOLOGY
N. Kaplan, Environmental Protection Agency, Research Triangle Park, North Carolina 387

OPERATING EXPERIENCE--CEA/ADL DUAL ALKALI PROTOTYPE SYSTEM AT GULF POWER/SOUTHERN SERVICES, INC.
C. R. LaMantia and R. R. Lunt, Arthur D. Little, Inc., Cambridge, Massachusetts; R. E. Rush, Southern Services, Inc., Birmingham, Alabama; T. M. Frank, Combustion Equipment Associates, Inc., New York, New York; and N. Kaplan, Environmental Protection Agency, Research Triangle Park, North Carolina 423

TITLE PAGE

THE FMC CONCENTRATED DOUBLE-ALKALI PROCESS
L. K. Legatski, K. E. Johnson, and L. Y. Lee,
FMC Corporation, Glen Ellyn, Illinois 471

OPERATING EXPERIENCE WITH THE ZURN DOUBLE ALKALI FLUE GAS DESULFURIZATION PROCESS
P. M. Lewis, Zurn Industries, Inc., Birmingham,
Alabama 503

KUREHA FLUE GAS DESULFURIZATION "SODIUM ACETATE-GYPSUM PROCESS"
S. Saito, T. Morita, and S. Suzuki, Kureha Chemical
Industry Company, Ltd., Tokyo, Japan 515

THE BUELL DOUBLE-ALKALI SO_2 CONTROL PROCESS
H. E. Bloss, Buell-Envirotech, Lebanon, Pennsylvania;
J. Wilhelm, EIMCO-Envirotech, Salt Lake City, Utah;
and W. J. Holhut, Central Illinois Public Service
Company, Springfield, Illinois 545

VOLUME II

BY-PRODUCT DISPOSAL/UTILIZATION SESSION xi

STATUS AND PLANS FOR WASTE DISPOSAL FROM UTILITY APPLICATIONS OF FLUE GAS DESULFURIZATION SYSTEMS
J. L. Crowe, Tennessee Valley Authority, Chattanooga,
Tennessee; and H. W. Elder, Tennessee Valley Authority,
Muscle Shoals, Alabama 565

RESEARCH AND DEVELOPMENT FOR CONTROL OF WASTE AND WATER POLLUTION FROM FLUE GAS CLEANING SYSTEMS
J. W. Jones, Environmental Protection Agency,
Research Triangle Park, North Carolina 579

FLUE GAS CLEANING WASTE DISPOSAL - EPA SHAWNEE FIELD EVALUATION
J. Rossoff and R. C. Rossi, The Aerospace Corporation,
El Segundo, California 605

CHEMICAL FIXATION OF FGD SLUDGES - PHYSICAL AND CHEMICAL PROPERTIES
J. L. Mahloch, U. S. Army Engineer Waterways Experiment
Station, Vicksburg, Mississippi 627

TITLE PAGE

POTENTIAL UTILIZATION OF CONTROLLED SO_x EMISSIONS FROM POWER PLANTS IN EASTERN UNITED STATES
J. I. Bucy, J. L. Nevins, P. A. Corrigan, and A. G. Melicks, Tennessee Valley Authority, Muscle Shoals, Alabama 647

REGENERABLE PROCESSES SESSION 701

STATUS OF DEMONSTRATION OF THE WELLMAN-LORD/ALLIED FGD SYSTEM AT NIPSCO'S D. H. MITCHELL GENERATING STATION
PART I: BACKGROUND AND OVERVIEW
E. L. Mann, Northern Indiana Public Service Company, Michigan City, Indiana; and R. C. Christman, TRW Environmental Engineering Division, Vienna, Virginia ... 703
PART II: CURRENT STATUS AND OPERATING PLAN
S. F. Lakatos, A. W. Michener, Jr., and W. D. Hunter, Jr., Allied Chemical Corporation, Morristown, New Jersey 709

AN UPDATE OF THE WELLMAN-LORD FLUE GAS DESULFURIZATION PROCESS
R. I. Pedroso, Davy Powergas, Inc., Lakeland, Florida .. 719

SUMMARY OF OPERATIONS OF THE CHEMICO-BASIC MgO FGD SYSTEM AT THE PEPCO DICKERSON GENERATING STATION
R. B. Taylor and P. R. Gambarani, Chemico Air Pollution Control Co., New York, New York; and D. Erdman, Potomac Electric Power Company, Washington, D. C. 735

MAGNESIUM OXIDE SCRUBBING AT PHILADELPHIA ELECTRIC'S EDDYSTONE STATION
J. A. Gille, Philadelphia Electric Company, Philadelphia, Pennsylvania 749

INTERIM REPORT ON CHIYODA THOROUGHBRED 101 COAL APPLICATION PLANT AT GULF POWER'S SCHOLZ PLANT
R. B. Dakan, Chiyoda International Corporation, Seattle, Washington; and R. A. Edwards and R. E. Rush, Southern Services, Inc., Birmingham, Alabama 761

ADVANCED PROCESSES 785

TITLE PAGE

STATUS AND ECONOMICS OF THE ATOMICS INTERNATIONAL AQUEOUS CARBONATE FLUE GAS DESULFURIZATION PROCESS
D. C. Gehri and R. D. Oldenkamp, Atomics International, Canoga Park, California 787

ENERGY REQUIREMENTS FOR SHELL FGD PROCESS
F. A. Vicari and J. B. Pohlenz, UOP Inc., Des Plaines, Illinois 817

THE DOWA'S BASIC ALUMINUM SULFATE-GYPSUM FLUE GAS DESULFURIZATION PROCESS
Y. Yamamichi and J. Nagao, The Dowa Mining Co., Ltd., Okayama, Japan 833

CITRATE PROCESS FOR FLUE GAS DESULFURIZATION, A STATUS REPORT
W. I. Nissen, D. A. Elkins, and W. A. McKinney, Bureau of Mines, Salt Lake City, Utah 843

APCI/IFP REGENERATIVE FGD AMMONIA SCRUBBING PROCESS
C. E. Ennis, Catalytic, Inc., Philadelphia, Pennsylvania 865

BF DRY ADSORPTION SYSTEM 877
PART I: FW-BF GULF POWER DEMONSTRATION UNIT INTERIM RESULTS
J. Strum and W. F. Bischoff, Foster Wheeler Energy Corporation, Livingston, New Jersey; and R. E. Rush, Southern Services, Inc., Birmingham, Alabama 879
PART II: BF-STEAG DEMONSTRATION UNIT OPERATIONAL EXPERIENCE AND PERFORMANCE
K. Knoblauch, Bergbau-Forschung GMBH, West Germany; and K. Goldschmidt, STEAG Aktengesellschaft, West Germany 899

THE CONSOL FGD PROCESS
R. T. Struck, E. Gorin, and W. E. Clark, Conoco Coal Development Company, Library, Pennsylvania 913

UNPRESENTED PAPERS 931

INFORMATION TRANSFER PROGRAM
T. W. Devitt and T. C. Ponder, Jr., PEDCo-Environmental Specialists, Inc., Cincinnati, Ohio 933

TITLE PAGE

CATALYTIC/WESTVACO DESULFURIZATION PROCESS PROTOTYPE DEMONSTRATION PROGRAM
Catalytic, Inc., Philadelphia, Pennsylvania 945

THE SPRING-NOBEL HOECHST PROCESS FOR SULFUR DIOXIDE RECOVERY FROM STACK GASES
W. H. Stark, A. A. Syme, and J. C. H. Chu, Spring Chemicals, Ltd., Toronto, Canada 981

STARTUP OF AMERICAN AIR FILTER'S SULFUR DIOXIDE REMOVAL SYSTEM AT THE KENTUCKY UTILITIES COMPANY'S GREEN RIVER STATION
A. H. Berst, American Air Filter Co., Inc., Louisville, Kentucky; and J. Reisinger, Kentucky Utilities Co., Central City, Kentucky 991

TCA SPHERE DEVELOPMENT AND EVALUATION
P. Sorenson, N. E. Takvoryan, and R. J. Jaworowski, UOP, Inc., Darien, Connecticut 999

OPENING SESSION

Chairman: Richard D. Stern
Chief, Process Technology Branch
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina

KEYNOTE ADDRESS

SULFUR OXIDE CONTROL AND ELECTRICITY PRODUCTION

Russell E. Train, Administrator

Environmental Protection Agency
401 M Street, S.W.
Washington, D.C. 20460

I appreciate the opportunity to present some of my thoughts regarding sulfur oxide control as it relates to electricity production. As you know, the legislative driving force for air pollution clean-up efforts is the Clean Air Act, as amended in 1970. This Act has had, and will continue to have, significant impact on the burning of fossil fuels in electric generating facilities.

The Act called for the establishment of national ambient air quality standards which set limitations on the ambient concentrations of various pollutants--SO_x, NO_x, CO, oxidants, and particulates--to minimize adverse effects on human health and welfare.

In order to achieve the ambient standards, the Act called for the development of State Implementation Plans (SIPs) which identify abatement measures to be taken within the several regions of each state in order to meet the ambient standards. SIPs may call for, among other things, restrictions on the emission of pollutants from various classes of facilities. These implementation plans are tailored to the particular air quality problems of each state, and the emission standards that they call for can be more or less stringent in different parts of the country. Working together, EPA and the states have now completed the development of these implementation plans. The Act also set a rather stringent timetable for the achievement of ambient standards; in most cases, such standards were to be met by mid-1975.

Further, the Act calls for the New Source Performance Standards (NSPS), which would require all new industrial sources to install the best demonstrated pollution control technology regardless of the source's location. This provision is designed to minimize the shift of new industrial facilities into regions of the country where emission control requirements are less stringent. It is interesting to note that, in some states, the emission limits set under the SIP's are more strict than the emission limits set by the NSPS.

How does all this affect the electrical utilities? How do utilities relate to the Clean Air Act and sulfur oxide emissions? Using the fossil fuels available to them -- coal, oil and natural gas -- power plants have produced electricity in vast amounts. Unfortunately, they have also produced vast amounts of an undesirable byproduct -- sulfur oxides.

As you well know, sulfur oxides and their derivatives present a serious threat to human health. It is an established fact that serious respiratory disease and illness--and even death--result from human exposure to these compounds. In turn, such health effects result in loss of work and pay by those so affected. The Congress declared that, in setting clean air standards to protect human health--such as the primary sulfur oxide standards--cost and difficulty of control were not to be issues in establishing either the standards themselves or the compliance schedules with which to meet them.

In 1971, the Environmental Protection Agency carried out the Congressional mandate and promulgated primary (health-related) and secondary (welfare-related) standards for the regulated air pollutants, including sulfur oxides. With approximately 970 fossil-fueled power plants in this country emitting 17 million tons or roughly 60 percent of all sulfur oxides in 1972, State Implementation Plans focused on those plants that were in the more highly polluted regions. The SIPs called for varying degrees of SO_x emission control for each plant--ranging from no control to control levels more stringent than the New Source Performance Standards.

The EPA has consistently represented that the most reasonable approach to meeting SO_2 ambient standards is by means of an emission control approach rather than by dilution methods. We believe this approach is particularly appropriate since recent data indicate that the oxidation products of SO_2, i.e., acidic sulfates, may cause more serious health problems than SO_2 itself.

At the present time, we do not have a detailed understanding of the mechanism by which SO_2 is converted to sulfates; however, it appears to involve the presence of other pollutants in the atmosphere and can occur over distances greater than 100 Km.

The problem of the regulation of SO_2 emissions to protect human health, in the light of this new evidence, was recently thoroughly reviewed by the National Academy of Sciences. The Academy noted significant deficiences in information available, both on the conversion of SO_2 to sulfates and on the health effects associated with various sulfur components in the atmosphere. Nevertheless, the Academy found the evidence to be persuasive enough to justify minimizing or preventing increases in emissions of SO_2.

EPA has a number of studies underway to shed more light on the atmospheric transformation of SO_2 to sulfates and on the health threat of sulfates. The Agency is now undertaking a five-year research program to significantly reduce uncertainty in our understanding of the sulfate problem.

Much has happened over the five years since the enactment of the Clean Air amendments. We have had an energy crisis that has caused us to take a long, hard look at where we get our energy sources and how we utilize that energy. When the Clean Air Act amendments were first enacted most people assumed that our supply of fossil fuels would remain abundant and that the problems of high sulfur emissions could be solved by burning low sulfur fuels, cleaning high sulfur fuels or by a combination of the two. Before the Arab embargo, many plants had made costly conversions to allow the burning of cheap oil and further take advantage of available lower sulfur content. Now all that is changed.

The embargo is over for now and imported oil is flowing into the country at nearly the rate it was before the embargo. We have learned of the dangers of becoming too dependent on foreign sources of fuel and that, increasingly, we must rely upon our own resources and our own technology. We have also learned that attainment of ambient air quality standards by mid-1975 was infeasible in many instances.

We are in the fortunate position of having within our national borders sixty percent of the world's coal. It seems only natural that we should seek to take advantage of this fact. But too many eyes are looking west to the low sulfur variety that can be strip mined, ignoring both the environmental consequences of such stripping and the ready availability of abundant high sulfur coals in the East. As you all do, EPA recognizes that under current conditions, supplies of low sulfur fuels are insufficient to go around, that some give and take is going to have to take place if we are going to have both sufficient power and air that is safe to breath. That is why new legislative proposals were submitted by the President to the Congress during February, 1975, as limited amendments to the 1970 Clean Air Act.

While we continue to feel that the Clean Air Act on the whole is a reasonable and needed response to the severe problems of air pollution, we also recognize that it perhaps lacks a certain flexibility. In addition, we feel that the President's proposals reflect a realistic concern for social impacts and cost effectiveness. Rather than weakening the air quality standards as so many have feared, in fact, the proposed amendments strengthen the Act by injecting a more common sense and defensible approach toward achieving those standards.

Let me summarize some of the most important elements of the proposed amendments to the Clean Air Act relative to stationary sources.

Compliance Schedule Extensions - this amendment would authorize compliance schedule extensions for certain isolated plants until January 1, 1985, at the latest, to install and operate scrubber systems or acquire low sulfur coal contracts. Under the compliance schedule, these plants could temporarily employ intermittent control systems if it could be demonstrated that they are reliable and enforceable. EPA estimates that there are approximately 18 to 70 plants which could use intermittent control prior to installation of permanent controls.

Waiver for Technology Innovations - this amendment would authorize, where the EPA Administrator approves in advance, a waiver of compliance with NSPS to encourage innovative and experimental control technology, provided primary air quality standards are met. This amendment would have important implications for second generation FGD and coal cleaning processes.

Assessment of Civil Penalties - This amendment proposes providing civil penalties of up to $25,000 per day of violation by stationary sources.

Enforcement Order Extensions - this change would clarify the Administrator's authority to issue enforceable compliance orders which extend beyond the statutory air quality standards attainment date.

Prevention of Significant Deterioration - EPA did not propose major modifications; other Federal agencies suggested that the total provision be deleted from the act.

Primary Particulate Standard - this amendment would allow the EPA Administrator to grant certain regions of the country additional time to meet the primary air standards for particulate matter.

New Source and Hazardous Emission Standards - this change explicitly authorizes EPA to set design or equipment standards for certain pollutant/source combinations if, and for as long as, it is not feasible to set customary emission or performance standards.

I believe these amendments would provide the Environmental Protection Agency with the flexibility needed to achieve compliance with air quality standards without weakening the basic thrust of the statute. I have been insistent that, in developing these needed changes, we in no way undermine the fundamental policies and objectives of the act.

The Congress has acted deliberately in considering these proposals and I expect that within the next three months we will see legislation generated on this subject. It is too early to say what the particulars of the final legislation will be, or how closely they will reflect EPA's recommendations. However, based on Congressional deliberations to date, I do not expect any significant weakening in the spirit of the Clean Air Act.

As you are well aware, EPA feels that the best available technology for controlling sulfur oxides when low-sulfur fuels are unavailable are the flue gas desulfurization systems which are the subject of this Symposium. It would be presumptuous of me to lecture a large body of experts on flue gas desulfurization on the viability and competitivenss of such processes. However, I will say that the Agency is keeping abreast of alternative SO_x emission control technologies such as physical and chemical coal cleaning, fluidized bed combustion, solvent refined coal and low-BTU coal gasification. It is clear to us that, with the exception of physical coal cleaning which can allow achievement of emission standards in a limited number of circumstances, only flue gas desulfurization presents a technology which can achieve required emission limitations consistent with the time frame mandated by the present or proposed changes to the Clean Air Act. Chemical coal cleaning, fluidized bed combustion, solvent refined coal and low-BTU gasification will be potentially competitive with FGD no earlier than the early-1980's.

We recognize that FGD technology is not yet a mature technology with all problems resolved to everyone's satisfaction. However, we believe this technology can and is playing an important role in minimizing SO_x emissions from utility and large industrial sources. I strongly support utility and government efforts to utilize and improve FGD technology; I am proud of the Office of Research and Development's Industrial Environmental Research Laboratory's sponsorship of FGD Symposia to communicate recent developments in this fast-moving technology. There is no question that further effort can lower costs, and produce more desirable end products. I trust that with mature cooperation from the industrial and government sectors we can optimize and further utilize this most important control technology.

At this time, 45 different utilities, General Motors, and Air Force and TVA have all committed to the use of FGD technology for one or more of their fossil fuel plants. Large and small utilities alike are represented. We find their commitment to cleaner air as well as to plentiful power admirable, especially when we consider that the capital and operating cost increments are not insignificant. I feel strongly that those utilities that lack the incentive to develop and install effective sulfur oxide control equipment and that expend great quantities of their resources to avoid such control measures will in the end have done a grave disservice to the public and to themselves.

Common sense dictates that the utilities must provide the power to keep America running and common sense tells us that they must use all of the energy sources that are available to them, including high sulfur coal. And I can assure you that the American people will not tolerate either a diminution of necessary electrical power or a deterioration of their health and general welfare caused by generation of that power. It should be clear to all that we cannot afford either of these alternatives. Adequate power is achievable. Clean air is obtainable.

REMARKS

Dr. Stephen Gage, Deputy Assistant Administrator

Office of Energy, Minerals, and Industry
Environmental Protection Agency
401 M Street, S.W.
Washington, D.C. 20460

Before I hand the microphone over to Dick Stern who will make some opening comments and introduce the first speaker, I would like to add a few remarks of my own. First, I should say that, in at least two instances, the Congressional committees considering the Clean Air Act amendments are taking a harder line than that embodied in the Administration's proposals. On the "significant deterioration" issue, it appears that the Congressional position is one of requiring quite stringent protection of those areas of the country where air quality is currently excellent. Also, the Congressional committees appear to be moving toward a position of requiring Best Available Control Technology (BACT) for all new plants, regardless of the quality of the fuel consumed. However, it is too early to tell what changes will be made in the House committee considering the bill, or what changes will be made on the floor in both houses, and on the conference committee.

Those of you who attended the last EPA-sponsored FGD symposium during November, 1974 in Atlanta may remember some of my comments as keynote speaker. One of the points I made then was that alternative technologies to flue gas desulfurization, particularly low-BTU gasification and fluidized bed combustion, cannot make commercial impact before the 1980's. Unfortunately, nothing has happened in the last year and a half to change my assessment.

As Mr. Train indicated in his keynote address, flue gas desulfurization represents the only major alternative to the burning of scarce, low-sulfur fuels in the next 8-10 years. However, I would like to mention one development which I believe has the potential to make an important near-term impact on SO_x control from coal burning from plants.

General Public Utilities Corporation (GPU) will install a physical coal cleaning plant at their 650 megawatt plant in Homer-City, Pennsylvania. This coal cleaning process utilizes a two-stage coal washing process of coal; one stream will have low sulfur content (0.8% or less) and the second stream will contain about 2.2% sulfur. Coal from both streams will be suitable for direct firing in the Homer-City boilers (both new and existing units) without further cleaning to meet state emission standards. EPA plans to participate in this important demonstration program. Such an approach can in certain cases achieve SIP standards or NSPS alone, or more generally can be used in conjunction with FGD in what may be the most cost-effective approach for many applications.

At the previous Symposium, I also discussed three Research, Development, and Demonstration initiatives that EPA's ORD planned, to further advance the state-of-the-art of FGD. First, I indicated that ORD would conduct an advanced lime and limestone test program on our flexible test facility at TVA's Shawnee plant. I am happy to report that this advanced program was initiated and has provided many important insights into lime and limestone scrubbing. As you will hear in two papers this afternoon: Operating conditions and mist eliminator configurations capable of long term operation at high velocities in a closed loop mode of operation have been identified. Operating conditions have been identified which have shown enhanced limestone utilization (less sludge production). Three commercial sludge fixation processes are being evaluated at a prototype scale.

Second, to provide an alternative to lime/limestone throwaway systems, I indicated that ORD would initiate a full-scale demonstration of a Double Alkali system. Although the evaluation process has moved more slowly than planned, I am happy to state that we have received excellent proposals and are in the final stages of selecting utility/vendor team for final negotiations. It is interesting to note that the total costs for each of the three proposals ranged from $60-75/Kw. Such systems seem to represent a promising alternative to lime scrubbing systems.

Third, to provide an alternative to sulfur-producing synthetic fuel processes, I indicated that ORD would initiate a full-scale demonstration of a regenerable FGD system which produces by-product sulfur. Again, the selection process has moved slowly, but we have received several viable proposals and are quite close to making our final selection. The selected system will be demonstrated at a 100 Mwe coal-fired boiler burning high-sulfur coal.

We hope to make our selections for both demonstrations public within a few weeks and trust that these programs will further advance the state-of-the-art of FGD technology.

Finally, I should mention that, in the past year, The Office of Research and Development has signed a Memorandum of Understanding with the Electric Power Research Institute in order to facilitate sharing of technical information and cooperation on Research, Development, and Demonstration projects. We are very hopeful that this cooperative research agreement will provide the basis for a creative approach to solving the remaining pollution control problems facing the electric power industry.

It has been my pleasure to appear before you today to present Mr. Train's remarks and make some comments of my own. Obviously, the Symposium is designed to be another success in the series of successful FGD symposia. The Industrial Environmental Research Laboratory is to be congratulated for sponsoring this series and for contributing so much to the development of FGD technology.

STATUS OF FLUE GAS DESULFURIZATION SYSTEMS IN THE UNITED STATES

Timothy W. Devitt, Dr. Gerald A. Isaacs,
and Bernard A. Laseke

PEDCo-Environmental Specialists, Inc.
Suite 13, Atkinson Square
Cincinnati, Ohio 45246

ABSTRACT

PEDCo conducts a bimonthly survey for the Industrial Environmental Research Laboratory/RTP and the Division of Stationary Source Enforcement of EPA on the status of flue gas desulfurization (FGD) systems. This paper presents an overview of the results of this survey including data on the number and total equivalent magawatts of FGD systems installed, under construction, and planned for U.S. utility application. Information is also presented on trends in process application and on the reliability of operating systems.

Approximately 109 FGD systems with a total equivalent rating of 42,128 MW are either operational, under construction, or planned. SO_2 removal efficiencies are generally in the range of 80 to 90 percent. Operability or reliability of several of the more recent systems is approximately 90 percent although many of the older, first generation systems continue to operate with lower values.

Preceding page blank

STATUS OF FLUE GAS DESULFURIZATION TECHNOLOGY IN THE UNITED STATES

In March 1974, PEDCo-Environmental was awarded a study by the U.S. Environmental Protection Agency to evaluate the status of flue gas desulfurization (FGD) technology in the United States. This study has two outputs. The first is a series of reports on individual FGD facilities. These reports present values for pertinent process designs and operating parameters and describe the operability of the unit from start-up to the time of the plant survey. To date 13 installations have been visited and another 16 will be inspected by July 1976.

The second study output is a series of bimonthly reports listing the systems that are operational, under construction or planned, and describing the operability of the existing systems during the preceeding two month period.

As of January 1976, the following flue gas systems were either operational or under construction in the United States: limestone, lime, activated carbon, double alkali, Wellman-Lord/Allied Chemical, Chiyoda Thorobred 101, sodium carbonate, and magnesium oxide. In addition, there are several other systems in the pilot plant stage.

As shown in Table 1, as of January 1976 there were 21 operational FGD systems on utility-size boilers, 20 units under construction and 67 units planned. The planned units include those where a contract has been awarded or a letter of intent has been signed and units for which the utility is presently requesting or evaluating bids.

Table 1. NUMBER AND TOTAL MW OF FGD SYSTEMS

Status	No. of units	MW
Operational	21	3,796
Under construction	20	7,026
Planned		
Contract awarded	10	3,761
Letter of intent	10	3,911
Requesting/evaluating bids	7	3,837
Considering only FGD systems	40	19,797
Total	108	42,128

By late 1976, approximately 10,000 MW of FGD capacity will be installed of which 40 percent will be retrofit applications and 60 percent will be on new power plants. By 1980, approximately 40,000 MW of FGD capacity will be installed of which 35 and 65 percent will be retrofit and new installations respectively. Figure 1 illustrates the growth of retrofit and new applications through 1980.

The different process applications in 1976 and 1980 are presented in Table 2. As is obvious from this table, the preponderance of units, both now and anticipated through 1980, is lime or limestone based systems. Currently, lime based systems account for 37 percent, on an equivalent magawatt basis, and will account for approximately 21 percent in 1980. Limestone based systems account for 44 percent in 1976 and will account for 30 percent in 1980. Regenerable processes will remain at approximately 6 to 8

FGD INSTALLATIONS

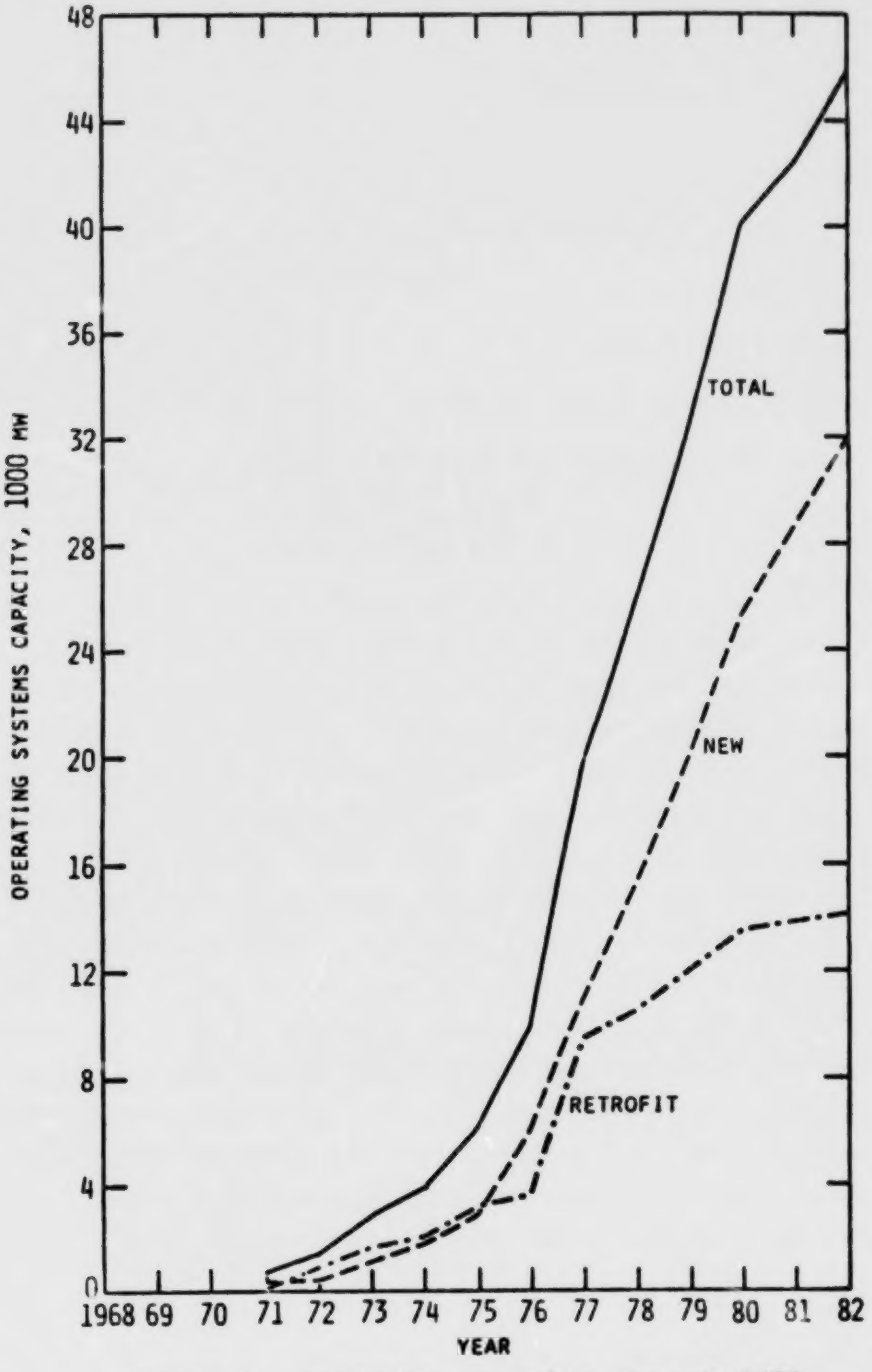


Figure 1. Operating capacity through 1982.

percent of the total during this period although the picture is clouded by the large number of planned systems for which the data is not available concerning process selection.

Table 2. FLUE GAS DESULFURIZATION PROCESSES

Process	1976	1980
Lime	3,442	7,716
Limestone	4,077	11,221
Lime/limestone	30	1,740
Activated carbon	20	20
Catox	110	110
Dual alkali	20	20
Magox	365	1,091
Sodium carbonate	375	375
Dilute acid	23	23
Wellman-Lord	115	1,830
Not selected	630	13,223
Total	9,207	37,369

SO_2 removal efficiencies for the operating units range from approximately 40 to 90 percent and particulate removal efficiencies generally are above 99 percent for those units designed for particulate removal. SO_2 removal efficiencies of approximately 98 percent have been attained during test runs at various facilities, although most systems are being designed to operate in the 80 to 90 percent efficiency range. FGD systems have been installed on units varying in size from about 30 MW to over 800 MW and on both low (0.4 to 1.0 percent) and high (6.0 percent) sulfur coal applications. The non-regenerable systems have used a variety of methods for sludge disposal including ponding, dewatering by vacuum filtration and landfilling, and fixation followed by landfilling. The regenerable systems have been limited to date to recovering sulfuric acid, although a sulfur recovery unit is being installed. Table 3 presents some of the values for key design and operating parameters for the operational FGD systems.

The only quantifiable process trend is that lime and limestone systems continue to predominate for all three FGD categories (i.e., operational, under construction, planned). Other apparent trends are:

(1) More systems are being installed to meet state standards that are more stringent than NSPS levels.

(2) More systems are being installed on low sulfur coal applications than on high sulfur coal applications.

(3) There is a tendency from wet fan applications to dry fan applications.

(4) Somewhat more attention is being paid to sparing of critical components (e.g., slurry pumps) although less than would be anticipated.

(5) There is a trend away from indirect reheat using tube bundles in the flue gas stream.

(6) Dry particulate collection systems (e.g., a high efficiency mechanical collector or an electrostatic precipitator) ahead of the FGD system are displacing particulate scrubber systems.

Table 3. DESIGN AND OPERATING DATA FOR SELECTED FGD SYSTEMS

Item No.	FGD process	Power Station		Size (MW)	No. modules	New/ retrofit	Start up date	By pass	Fuel Characteristics		
		Utility	Unit						Type	Btu/lb	% S
1	Lime scrubbing	Duquesne Light Co.	Elrama	510	5	Retrofit	10/75	yes	coal		1.0-2.8
2	Lime scrubbing	Duquesne Light Co.	Phillips	410	5	Retrofit	7/73	yes	coal	11,000	1.0-2.8
3	Lime scrubbing	Kentucky Utilities	Green River No. 1 & 2	64	2	Retrofit	9/75		coal	11,000	3.8
4	Lime scrubbing	Louisville Gas & Elec.	Paddy's Run	65	2	Retrofit	4/73	no	coal	11,500	3.5-4.0
5	Lime scrubbing	Montana Power Co.	Colstrip 1	360		New	10/75		coal	8,800	0.8
6	Lime scrubbing	S. Calif. Edison Co.	Mohave No. 2	160	1	Retrofit	11/73	yes	coal	11,500	0.4
7	Limestone scrubbing	Arizona Pu. Serv.	Cholla No. 1	115	2	Retrofit	10/73	yes	coal	12,146	0.44
8	Limestone scrubbing	Commonwealth Edison Co.	Will County No. 1	167	2	Retrofit	2/72	yes	coal	9,463	2.1
9	Limestone scrubbing	Kansas City Pwr & Light	Hawthorn No. 3	140	2	Retrofit	11/72	yes	coal	9,800-11,500	0.6-3.0
10	Limestone scrubbing	Kansas City Pwr & Light	Hawthorn No. 4	100	2	Retrofit	8/72	yes	coal	9,800-11,500	0.6-3.0
11	Limestone scrubbing	Kansas City Pwr & Light	La Cygne No. 1	820	7	New	2/73	no	coal	9,000	5.2
12	Limestone scrubbing	Kansas Pwr & Light	Lawrence No. 4	125	2	Retrofit	12/68	no	coal	10,000 12,000	0.6-3.75
13	Limestone scrubbing	Kansas Pwr & Light	Lawrence No. 5	400	8	New	11/71	yes	coal	10,000 12,000	0.6-3.75
14	Limestone scrubbing	Key West Utility Brd	Stock Island	37	2	New	10/72	yes	oil		2.4-2.75
15	Limestone scrubbing	Public Serv of Colorado	Valmont No. 5	50	1	Retrofit	10/74	yes	coal/ gas	10,780	0.72
16	Limestone scrubbing	S. Calif. Edison Co.	Mohave No. 1	170	1	Retrofit	11/74	yes	coal	11,500	0.5-0.8
17	Lime/Limestone scrubbing	Tennessee Valley Authority	Shawnee No. 10A	10	1	Retrofit	4/72	yes	coal		2.9
18	Lime/Limestone scrubbing	Tennessee Valley Authority	Shawnee No. 10B	10	1	Retrofit	4/72	yes	coal		2.9
19	Magnesium oxide scrubbing	Philadelphia Elec. Co.	Eddystone No. 1	120	1	Retrofit	9/75	yes	coal	12,100	2.3
20	Sodium carbonate scrubbing	Nevada Power Co.	Reid Gardner No. 1	125	1	Retrofit	4/74	yes	coal	12,450	0.5-1.0
21	Sodium carbonate scrubbing	Nevad. Power Co.	Reid Gardner No. 2	125	1	Retrofit	12/73	yes	coal	12,450	0.5-1.0
22	Double alkali	Gulf Power Co.	Scholz No. 1	20	1	Retrofit	2/75	yes	coal	12,400	3.0-5.0
23	Dilute acid	Gulf Power Co.	Scholz Nos. 1 & 2	23	1	Retrofit	3/75	yes	coal	12,400	3.0-5.0

Table 3 (continued). DESIGN AND OPERATING DATA FOR SELECTED FGD SYSTEMS

Item No.	Particulate control device	Control System Efficiency		Gas reheat		By-product		FGD System costs $/kW
		Particulate	SO_2	Type	Mode	Type	Disposal	
1						stabilized sludge	pond	
2	mech, esp, venturi	99	60	oil-fired	direct	stabilized sludge	haulaway	84
3	mech.	99.7	80			stabilized sludge	unlined pond	62.5
4	ESP			gas-fired	direct	unstabilized sludge	landfill	57
5		99	40			unstabilized sludge	unlined pond	50
6	ESP	N/A	75-98	hot air	indirect	stabilized sludge	lined pond	N/A
7	mech.	99.7	85	steam	indirect	unstabilized sludge	unlined pond	57
8	ESP, venturi	98	82	steam	indirect	stabilized sludge	clay lined pond	103
9	absorber tower	99	70	steam	indirect	unstabilized sludge	unlined pond	19
10	absorber tower	99	70	steam	indirect	unstabilized sludge	unlined pond	19
11	venturi	98	72	steam/air	indirect/direct	unstabilized sludge	unlined pond	43
12	absorber tower	99	75	steam	indirect	unstabilized sludge	unlined pond	N/A
13	absorber tower	99	65	steam	indirect	unstabilized sludge	unlined pond	N/A
14	mech.	90	85	none	none		settling pond	21.6
15	absorber tower	90	85			unstabilized sludge	lined pond	
16	ESP	N/A	N/A	steam	direct		converted to aggregate	N/A
17	absorber tower		75-88	fuel oil fired	direct		settling pond	N/A
18	venturi		80-98	fuel oil fired	direct		settling pond	N/A
19	mech., ESP, ventri-rod, scrubber	99.9	90	fuel oil fired	direct	magnesium sulfite	acid plant	193
20	mech., venturi	99	85	hot air	direct	spent liquor	settling pond	44
21	mech., venturi	99	85	hot air	direct	spent liquor	settling pond	44
22	ESP	99.7	N/A	fuel oil fired	direct	filter cake	lined pond	N/A
23	ESP, venturi	99.7	90+	fuel oil fired	direct	gypsum	lined pond	130

Several parameters have been suggested to quantify the viability of FGD systems. Various terms such as "availability," "reliability," and "operability" have been used indiscriminately in reference to these parameters so that a certain degree of confusion exists in defining these terms when they are used. These parameters are defined and briefly discussed below.

Availability (Parameter 1): Hours the FGD system was available for operation (whether operated or not) divided by hours in period, expressed as a percentage. This parameter tends to overassess the viability of the FGD system because it does not penalize for election not to operate the system when it could have been operated. Boiler downtime tends to increase the magnitude of the parameter because FGD failures are not included during such periods.

Operability (Parameter 2): Hours the FGD system was operated divided by boiler operating hours in the period, expressed as a percentage. This parameter indicates the degree to which the FGD system is actually used, relative to boiler operating time. The principal objection is that the parameter does not reflect the extent of exertion on the FGD system. That is, the magnitude of the parameter has little or no correlation with FGD system operating time. Another objection is that the parameter is penalized when options are exercised to not use the FGD system in periods when the system is operable.

Utilization Factor (Parameter 3): Hours that the FGD system operated divided by total hours in period. This parameter is a relative stress factor for the FGD system. It is not a complete measure of FGD system viability because the parameter can be strongly influenced by conditions that are external to the FGD system. For example, infrequent boiler operation will lower the value of the parameter although the FGD system may be highly dependable in its particular application.

Another parameter (Parameter 4), has been defined as the hours the FGD system was operated divided by the hours the FGD system was called upon to operate, expressed as a percentage. This parameter has been suggested in order not to penalize the FGD system for elected outages, e.g., periods when the FGD system could have been run but was not run because of chemical shortages, lack of manpower, short duration boiler operations, and the like. The main problem in using this formula is the concise determination whether or not the system was "called upon to operate" during a given time period.

Thus, no single ratio yields a completely desirable parameter to assess the degree of operability of an FGD system. It does seem that Parameters 2 and 3 together form a useful index of FGD system operability. Parameter 2 is a measure of responsiveness of the FGD system to the boiler's requirements, and Parameter 3 is a measure of the intensity at which the FGD system is operated. Perhaps "availability" or "reliability" should be quantified by two parameters instead of one, e.g., 78 percent operability/65 percent utilization. Figure 2 illustrates the difference in values for the various parameters using data from the experimental operation of the Mohave 170 MW horizontal unit FGD system from January 6, 1974 to February 9, 1975.

In most of our reports to date, we have used the operability parameter, FGD system operating hours over boiler operating hours, because the data required to calculate this parameter are the most readily available. The

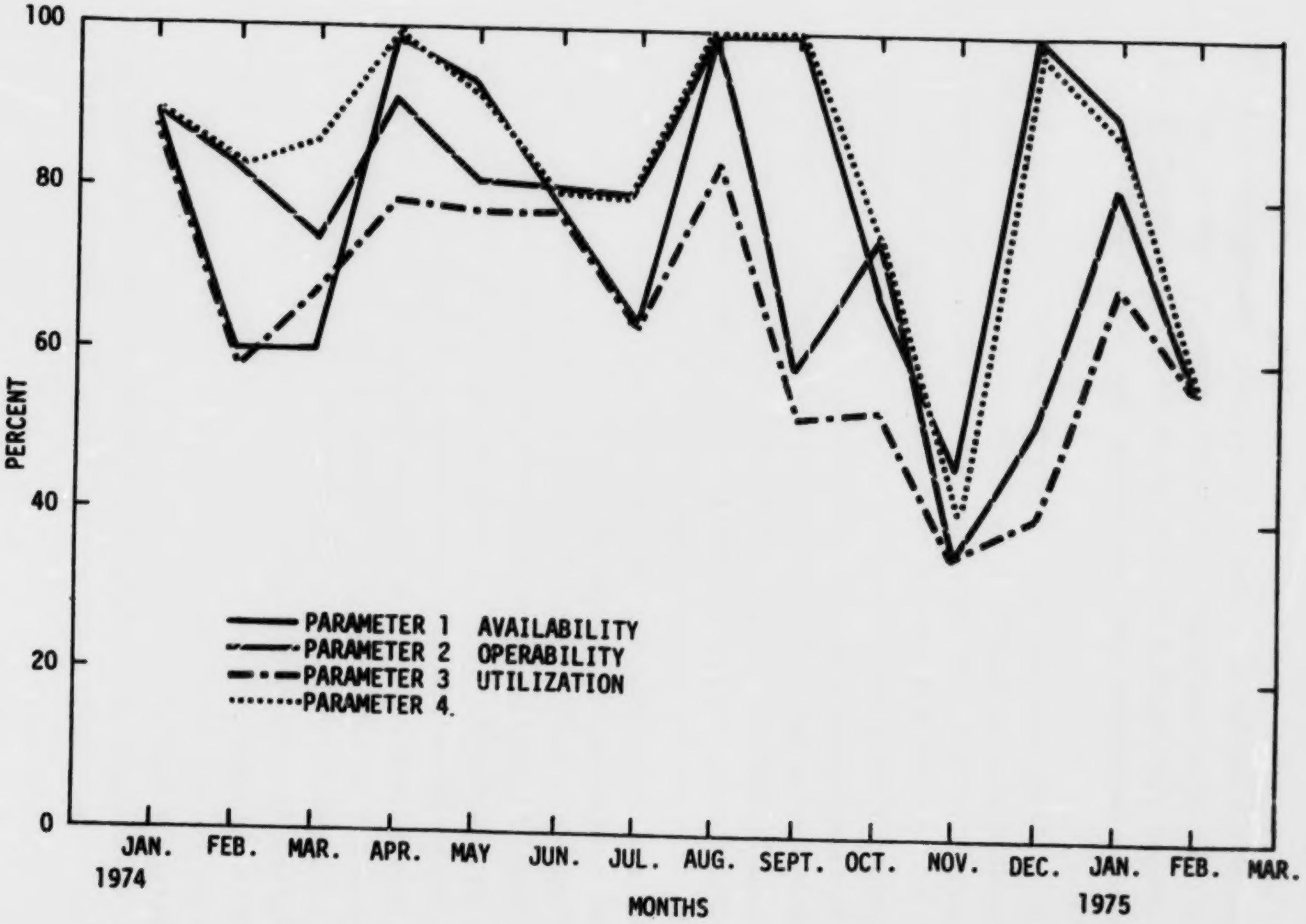


Figure 2. Comparison of FGD viability parameters for the Mojave Station of Southern California Edison.

operability of units has improved over the past 2 years as both the number of units and the operating experience has increased. Figures 3, 4 and 5 illustrate the operability of some of the more successful units by month of operation. The average operability of the units range from about 80 to 95 percent depending upon the system and the averaging period.

A utility industry survey was conducted recently by the Edison Electric Intitute to determine the projected operability/reliability and cost for FGD systems. Any utility known to have an FGD system that was operational, under construction or planned was contacted by EEI and requested to complete a questionnaire describing their FGD system. Responses were received for forty-three systems and were analyzed by EEI and PEDCo. For utilities with units under construction or in the planning stage, most estimated the reliability of the systems would be 90 percent or greater, which is essentially that being attained by some of the more recent operating units, as described above.

The cost side of the survey is also interesting. As anticipated, the reported costs cover a broad range due to both site-specific factors and the lack of uniformity with respect to items included in the cost estimates.

The responses covered 30 utilities, 32 plants and 68 boilers for a total capacity of 32,120 MW. The reported costs ranged from 33 to 197 $/kW with an average of $94/kW (standard deviation = $40/kW). Of these, 22 were lime or limestone based systems. The cost for these systems ranged from 34 to 116 $/kW with an average of $78/kW (standard deviation = $27/kW).

When these costs were adjusted to reflect a common year dollar value (1975), removal of costs for particulate emis-

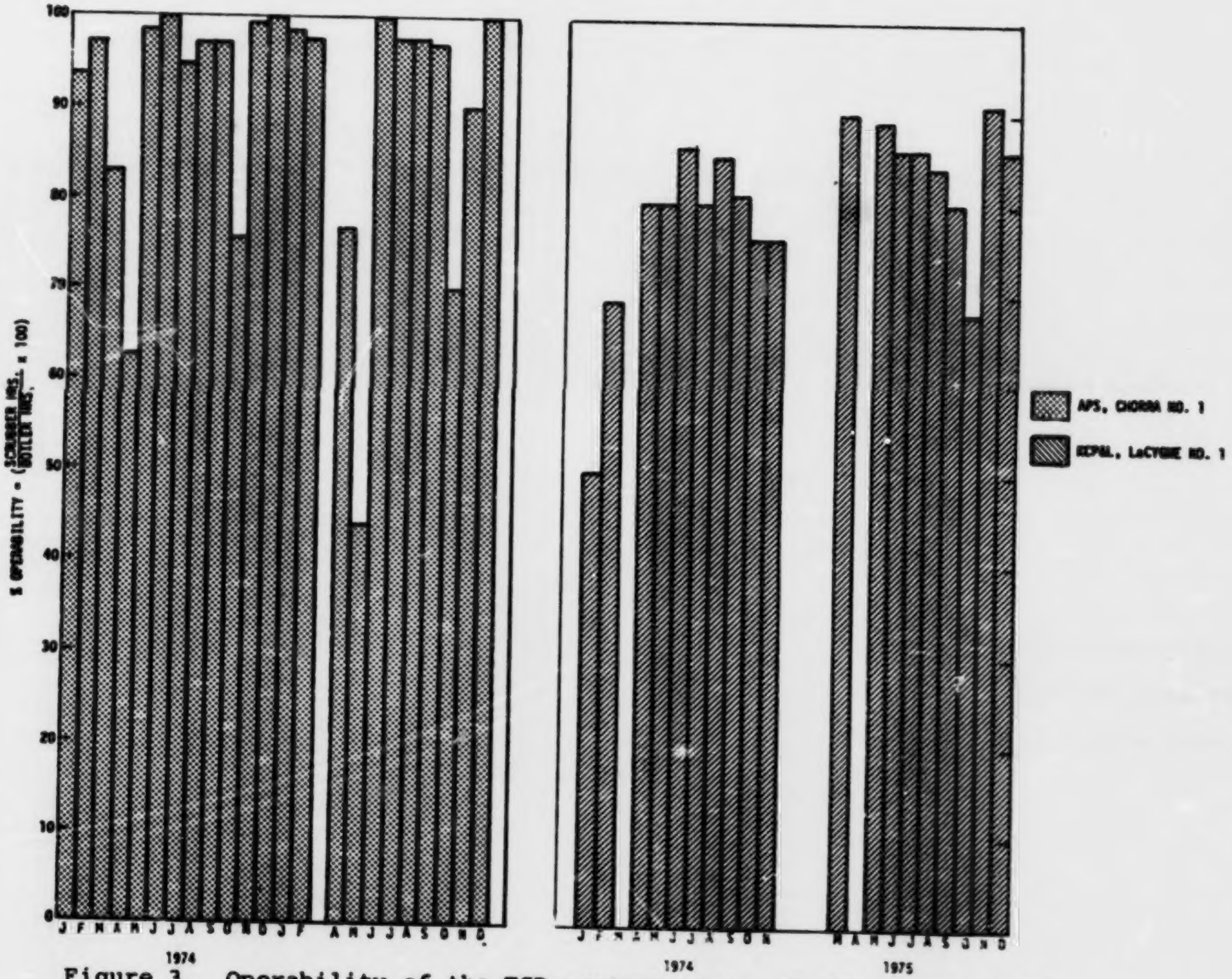


Figure 3. Operability of the FGD systems at the Cholla Power Plant, Arizona Public Service, and the LaCygne Station, Kansas City Power and Light.

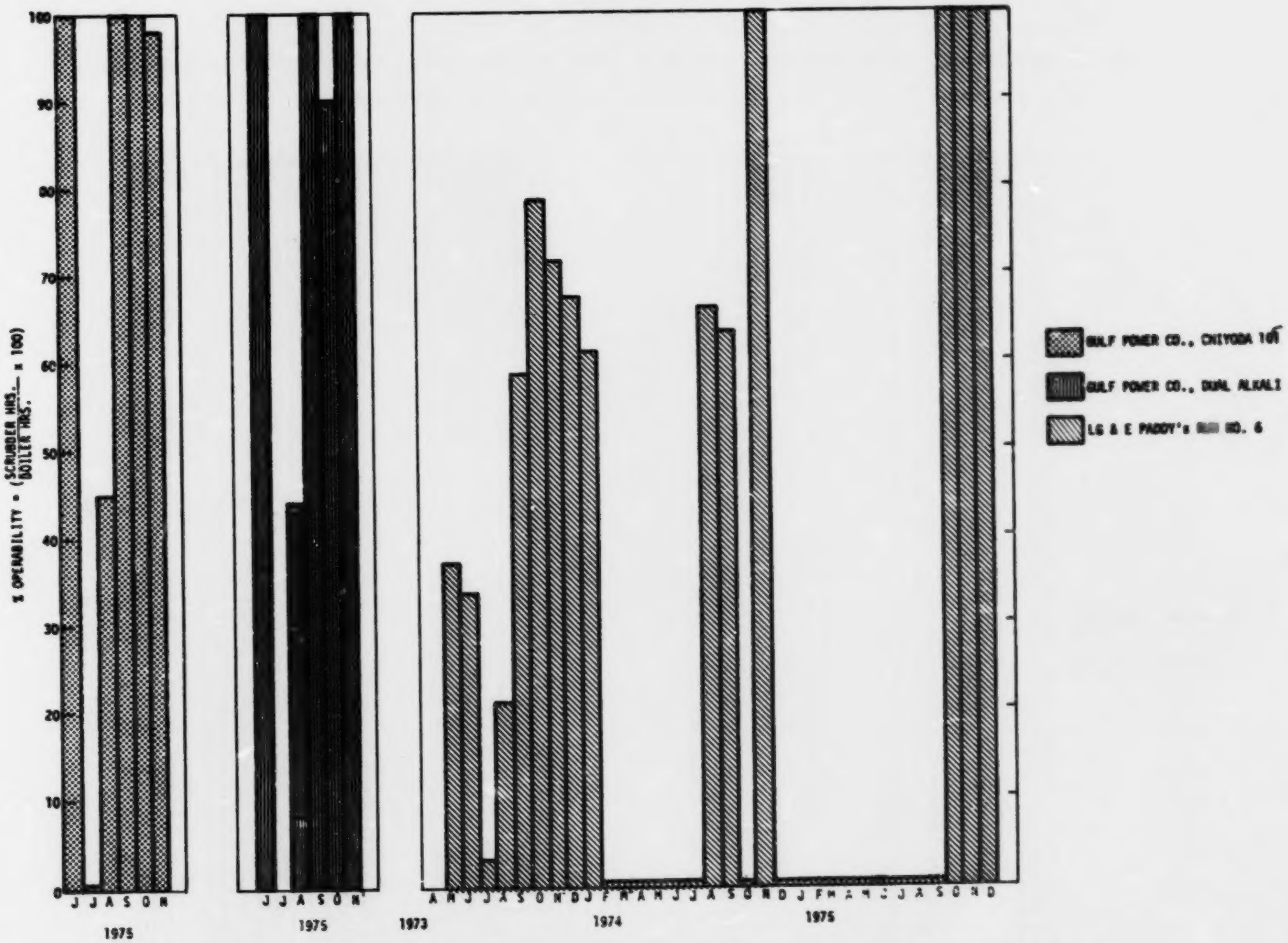


Figure 4. Operability of the FGD systems at the Scholz Station of Gulf Power Company and the Paddy's Run Station of Louisville Gas and Electric.

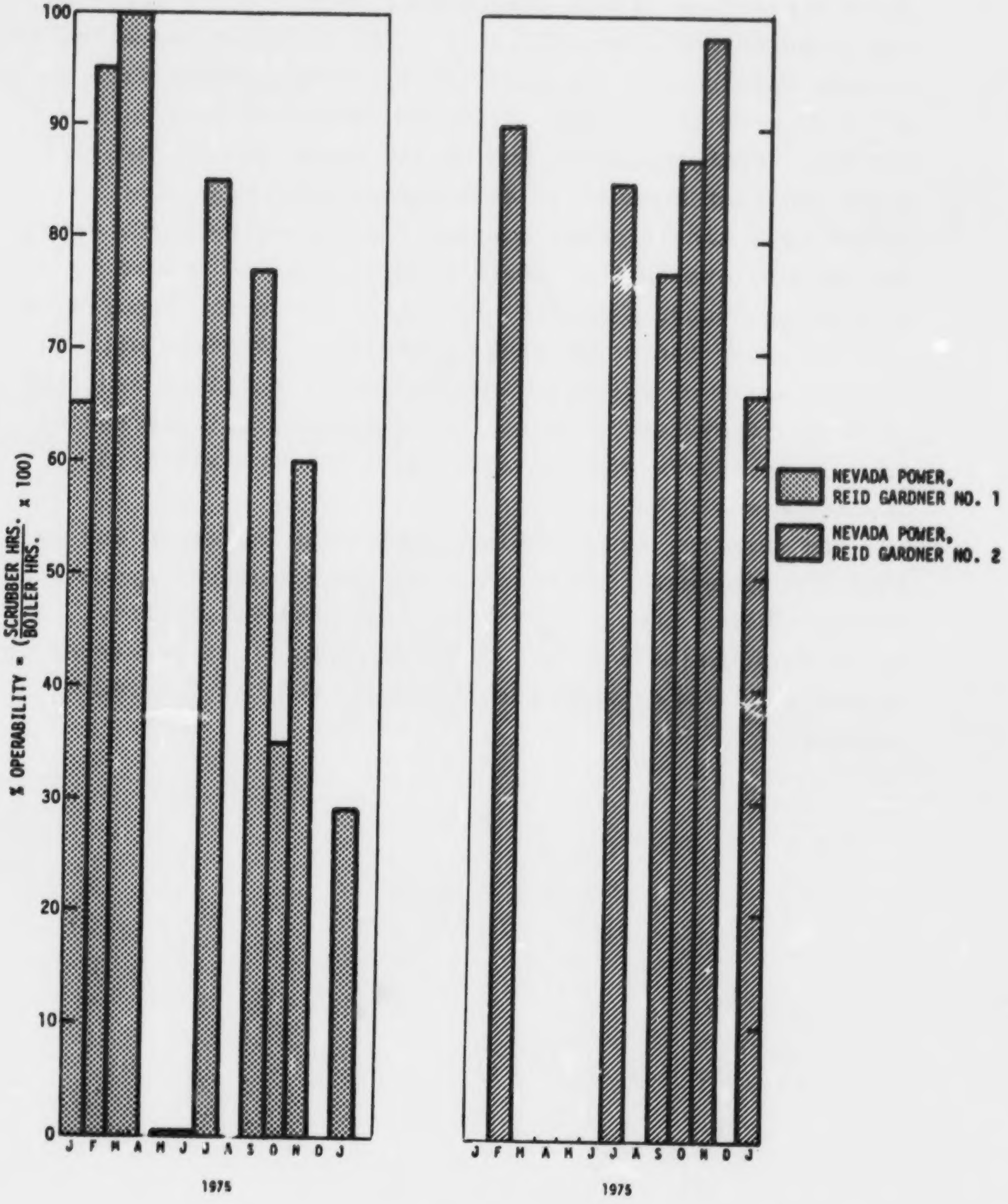


Figure 5. Operability of the FGD systems at the Reid Gardner Station of Nevada Power Company.

sion control, addition of indirect cost items and adequate costs for either sludge disposal or regeneration systems, the range narrowed considerably. The adjusted costs for all systems with sufficient data (30 systems), ranged from 50 to 205 $/kW with an average of $91/kW (standard deviation = $34/kW). Both the upper end of the range and the average costs are high because of an exceptionally high cost reported by a single power company for a prototype FGD system; the utility stated that their reported values should be considered "upper limits." Excluding the costs reported by that company, the costs range from 50 to 137 $/kW with an average value of $85/kW. Adjusted costs for lime and limestone based systems reported by nineteen utilities ranged from 50 to $88/kW with an average of $70/kW (standard deviation = $10/kW).

In summary, as operating experience increases and as this experience is fed back to the design phase, systems are coming on-line that can perform with high reliability. A table from the most recent bimonthly report is presented in Appendix A, summarizing the status of the various FGD systems.

APPENDIX A

STATUS OF FLUE GAS DESULFURIZATION SYSTEMS[a]

---

[a] Appendix A is Table 2 from the PEDCo November - December Bimonthly Report on the Status of Flue Gas Desulfurization Systems.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 1 ALABAMA ELECTRIC COOP TOMBIGBEE NO. 2 225 MW - NEW COAL 0.8- 1.5 PERCENT SULFUR PEABODY ENGINEERING LIMESTONE SCRUBBING STARTUP 3/78	PEABODY ENGINEERING WAS AWARDED CONTRACT TO INSTALL FGD SYSTEM ON TOMBIGBEE NO.2. SYSTEM WILL BE LIMESTONE SCRUBBING.
I.D. NUMBER 2 ALABAMA ELECTRIC COOP TOMBIGBEE NO. 3 225 MW - NEW COAL 0.8- 1.5 PERCENT SULFUR PEABODY ENGINEERING LIMESTONE SCRUBBING STARTUP 3/79	PEABODY ENGINEERING WAS AWARDED CONTRACT TO INSTALL FGD SYSTEM ON TOMBIGBEE NO.3. SYSTEM WILL BE LIMESTONE SCRUBBING.
I.D. NUMBER 3 ALLEGHENY POWER SYSTEM PLEASANTS NO. 1 625 MW - NEW COAL 2.0- 5.0 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 3/79	NEW STARTUP DATE IS SET FOR MARCH 1979. FGD SYSTEM IS IN THE PLANNING STAGE. ANTICIPATE DECISION ON THE CONTRACT SOON.
I.D. NUMBER 4 ALLEGHENY POWER SYSTEM PLEASANTS NO. 2 625 MW - NEW COAL 2.0- 5.0 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 3/80	NEW STARTUP DATE IS SET FOR MARCH 1980. FGD SYSTEM IS IN THE PLANNING STAGE. ANTICIPATE DECISION ON THE CONTRACT SOON.
I.D. NUMBER 5 ARIZONA ELECTRIC POWER COOP APACHE NO 2 200 MW - NEW COAL 0.5- 0.8 PERCENT SULFUR RESEARCH COTTRELL LIMESTONE SCRUBBING STARTUP 6/79	LETTER OF INTENT TO INSTALL AN FGD SYSTEM WAS ISSUED TO RESEARCH COTTRELL. ENGINEERING DESIGN WORK IS PROCEEDING.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 6 ARIZONA ELECTRIC POWER COOP APACHE NO 3 205 MW - NEW COAL 0.5- 0.8 PERCENT SULFUR RESEARCH COTTRELL LIMESTONE SCRUBBING STARTUP 6/79	LETTER OF INTENT TO INSTALL AN FGD SYSTEM WAS ISSUED TO RESEARCH COTTRELL. ENGINEERING DESIGN WORK IS PROCEEDING.
I.D. NUMBER 7 ARIZONA PUBLIC SERVICE CHOLLA NO 1 115 MW - RETROFIT COAL 0.44-1 PERCENT SULFUR RESEARCH COTTRELL LIMESTONE SCRUBBING STARTUP 10/73	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE NOVEMBER,1973. AVAILABILITY FOR MODULE A HAS AVERAGED OVER 95 PERCENT FOR THE PAST 6 MONTHS. AVAILABILITY FOR MODULE B HAS AVERAGED OVER 88 PERCENT FOR THE PAST 6 MONTHS. AVAILABILITY FOR MODULE A WAS 100 PERCENT IN NOVEMBER AND DECEMBER. AVAILABILITY FOR MODULE B WAS 80 PERCENT IN NOVEMBER AND 100 PERCENT IN DECEMBER.
I.D. NUMBER 8 ARIZONA PUBLIC SERVICE CHOLLA NO 2 250 MW - NEW COAL 0.44-1 PERCENT SULFUR RESEARCH COTTRELL LIMESTONE SCRUBBING STARTUP 6/77	RESEARCH COTTRELL WAS AWARDED A CONTRACT TO INSTALL FGD SYSTEM ON CHOLLA NO. 2, AND IS PRESENTLY WORKING ON THE DESIGN PHASE OF THIS PROJECT. ENGINEERING IS SUBSTANTIALLY COMPLETE. CONSTRUCTION IS UNDERWAY.
I.D. NUMBER 9 ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 4 755 MW - RETROFIT COAL, 0.7 - 0.75% SULFUR (AVG) NOT SELECTED NOT SELECTED STARTUP 0/ 0	THIS 755 MW UNIT WILL BE CONTROLLED PENDING THE OUTCOME OF TESTS TO BE MADE ON THE PROTOTYPE 160 MW FGD MODULE NOW BEING INSTALLED ON UNIT 5.
I.D. NUMBER 10 ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 5A 160 MW - RETROFIT COAL, 0.7 - 0.75, SULFUR (AVG) SCE LIME SCRUBBING STARTUP 11/75	THIS PROTOTYPE 160 MW SYSTEM IS BEING MOVED FROM THE MOHAVE STATION OF SOUTHERN CALIFORNIA EDISON, WHERE THE SYSTEM HAS BEEN TESTED PREVIOUSLY. IT WILL HANDLE ABOUT 20% OF THE EXHAUST GAS FROM UNIT 5 RATED AT 755MW (NET). ENGINEERING IS 80% COMPLETE AND CONSTRUCTION IS 55% COMPLETE. PARTICULATE AND SO2 EFFICIENCIES FOR THIS UNIT ARE TO BE DETERMINED, AND UNIT RELIABILITY WILL BE EVALUATED. IT HAS NOT YET BEEN DETERMINED IF THIS INSTALLATION WILL BE PERMANENT.

TABLE 2

STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 11 ARIZONA PUBLIC SERVICE FOUR CORNERS NO. 5B 595 MW - RETROFIT COAL, 0.7 - 0.75, SULFUR (AVG) NOT SELECTED NOT SELECTED STARTUP 0/ 0	THE BALANCE OF THIS 755 MW UNIT WILL BE CONTROLLED PENDING THE OUTCOME OF TESTS TO BE MADE ON THE PROTOTYPE 160 MW FGD MODULE NOW BEING INSTALLED NO START-UP DATE HAS BEEN SCHEDULED.
I.D. NUMBER 12 BASIN ELECTRIC POWER COOP MISSOURI BASIN NO 1 550 MW - NEW COAL NOT SELECTED REGENERABLE NOT SELECTED STARTUP 1/80	BECAUSE OF THE STRICT WYOMING EMISSION STANDARDS OF 0.2LBS PER MILLION BTU,LOW SULFUR COAL ALONE WILL NOT MEET STANDARDS. BASIN ELECTRIC IS NOW INVESTIGATING VARIOUS REGENERABLE PROCESSES.
I.D. NUMBER 13 BASIN ELECTRIC POWER COOP MISSOURI BASIN NO 2 550 MW - NEW COAL NOT SELECTED REGENERABLE NOT SELECTED STARTUP 6/80	BECAUSE OF THE STRICT WYOMING EMISSION STANDARDS OF 0.2LBS PER MILLION BTU,LOW SULFUR COAL ALONE WILL NOT MEET STANDARDS. BASIN ELECTRIC IS NOW INVESTIGATING VARIOUS REGENERABLE PROCESSES.
I.D. NUMBER 14 BASIN ELECTRIC POWER COOP MISSOURI BASIN NO 3 550 MW - NEW COAL NOT SELECTED REGENERABLE NOT SELECTED STARTUP 6/85	BECAUSE OF THE STRICT WYOMING EMISSION STANDARDS OF 0.2LBS PER MILLION BTU,LOW SULFUR COAL ALONE WILL NOT MEET STANDARDS. BASIN ELECTRIC IS NOW INVESTIGATING VARIOUS REGENERABLE PROCESSES.
I.D. NUMBER 15 CENTRAL ILLINOIS LIGHT CO. DUCK CREEK NO.1 100 MW - NEW COAL 2.5-3.0 PERCENT SULFUR RILEY STOKER/ENVIRONEERING LIMESTONE SCRUBBING STARTUP 6/76	THE UNIT IS NOW UNDER CONSTRUCTION AND IS SCHEDULED FOR COMPLETION IN APRIL 1976. STATION CAPACITY IS 400MW, BUT FLUE GAS FROM ONLY 100 MW WILL BE TREATED INITIALLY.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 16 CENTRAL ILLINOIS LIGHT CO. DUCK CREEK NO.2 400 MW - NEW COAL 2.5-3.0 PERCENT SULFUR NOT SELECTED LIMESTONE SCRUBBING STARTUP 1/81	THIS PROPOSED LIMESTONE UNIT IS DUE TO BEGIN OPERATION JANUARY 1981
I.D. NUMBER 17 CENTRAL ILLINOIS LIGHT CO. E.D.EDWARDS NO.3 357 MW - RETROFIT COAL 2.5 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 7/79	LOW SULFUR COAL TO BE USED FROM 6/76 UNTIL STARTUP.
I.D. NUMBER 18 CENTRAL ILLINOIS PUBLIC SERV NEWTON NO.1 600 MW - NEW COAL 2.8-3.2 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 12/77	ALL BIDS HAVE BEEN RECEIVED AND ARE NOW BEING EVALUATED. THE PROCESSES BEING CONSIDERED ARE LIME/LIMESTONE SCRUBBING AND DOUBLE ALKALI. A DECISION ON THE PROCESS AND VENDOR SHOULD BE MADE BY FEBRUARY 1,1976
I.D. NUMBER 19 CINCINNATI GAS & ELECTRIC CO. EAST BEND NO 2 600 MW - NEW COAL NOT SELECTED NOT SELECTED STARTUP 1/80	UTILITY WILL GO OUT FOR BIDS IN EARLY 1976. THE COMPANY IS PRESENTLY OBTAINING CONSTRUCTION AND OPERATIONG PERMITS. STARTUP DATE HAS BEEN MOVED BACK TO JANUARY 1980. THE COAL SOURCE FOR THIS UNIT HAS NOT YET BEEN SELECTED.
I.D. NUMBER 20 CINCINNATI GAS & ELECTRIC CO. MIAMI FORT NO 8 300 MW - NEW COAL 1.3 PERCENT SULFUR NOT SELECTED LIME/LIMESTONE SCRUBBING STARTUP 1/78	THE UTILITY HAS SIGNED A CONTRACT FOR THE PURCHASE OF LOW SULFUR COAL (APPROXIMATELY 1.3 PERCENT SULFUR). THE COAL BED FOR THE NO.8 UNIT IS NOW IN THE PROCESS OF BEING REDESIGNED TO ACCEPT THE LOWER SULFUR COAL. CG&E NOW PLANS TO INSTALL AN FGD UNIT TREATING 60 PERCENT OF THE TOTAL 500 MW CAPACITY. NEW BIDS ARE NOW BEING REQUESTED AND ACCEPTED ON THIS SMALLER UNIT.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 21 COLUMBUS & SOUTHERN OHIO ELEC. CONESVILLE NO 5 400 MW - NEW COAL 4.5 - 4.9 PERCENT SULFUR UNIVERSAL OIL PRODUCTS LIME SCRUBBING STARTUP 6/76	THE UTILITY SIGNED LONG-TERM CONTRACTS WITH DRAVO FOR THE PURCHASE OF THIOSORBIC LIME AND WITH IUCS FOR THE PURCHASE OF SLUDGE FIXATIVES. THE BOILER AND FGD SYSTEM ARE STILL UNDER CONSTRUCTION. THIS MINE MOUTH PLANT PLANS TO BURN COAL WITH 17 PERCENT ASH CONTENT AND 4.5 TO 4.9 PERCENT SULFUR CONTENT. AN ELECTROSTATIC PRECIPITATOR WILL BE INSTALLED UPSTREAM OF THE FGD SYSTEM.
I.D. NUMBER 22 COLUMBUS & SOUTHERN OHIO ELEC. CONESVILLE NO 6 400 MW - NEW COAL 4.5 - 4.9 PERCENT SULFUR UNIVERSAL OIL PRODUCTS LIME SCRUBBING STARTUP 1/78	THE UTILITY SIGNED LONG-TERM CONTRACTS WITH DRAVO FOR THE PURCHASE OF THIOSORBIC LIME AND WITH IUCS FOR THE PURCHASE OF SLUDGE FIXATIVES. CONSTRUCTION ON THIS UNIT IS CURRENTLY SCHEDULED FOR 1977 WITH COMPLETION EXPECTED BY JANUARY 1978. SIMILAR TO CONESVILLE NO.5, THIS MINE MOUTH PLANT WILL BURN COAL WITH 17 PERCENT ASH CONTENT AND 4.5 TO 4.9 PERCENT SULFUR CONTENT. AN ELECTROSTATIC PRECIPITATOR WILL BE INSTALLED UPSTREAM OF THE FGD SYSTEM.
I.D. NUMBER 23 COLUMBUS & SOUTHERN OHIO ELEC. POSTON NO. 5 375 MW - NEW COAL 2.5 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 0/81	UNIT WILL BURN 2.5 PERCENT SULFUR OHIO COAL.
I.D. NUMBER 24 COLUMBUS & SOUTHERN OHIO ELEC. POSTON NO. 6 375 MW - NEW COAL 2.5 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 0/83	UNIT WILL BURN 2.5 PERCENT SULFUR OHIO COAL.
I.D. NUMBER 25 COMMONWEALTH EDISON WILL COUNTY NO 1 167 MW - RETROFIT COAL - 9463 BTU, 2.1% SULFUR BABCOCK & WILCOX LIMESTONE SCRUBBING STARTUP 2/72	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE FEBRUARY,1972. AVAILABILITY FOR MODULES A AND B DURING THE 12 MONTH PERIOD FROM OCTOBER 1974 TO SEPTEMBER 1975 AVERAGED 54 AND 25 PERCENT RESPECTIVELY. BOTH MODULES WERE OUT OF SERVICE DURING THE REPORT PERIOD DUE TO A SCHEDULED OVERHAUL OF THE BOILER, TURBINE, AND SCRUBBER MODULES.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 26 DETROIT EDISON ST.CLAIR NO 6 180 MW - RETROFIT COAL 3.7 PERCENT SULFUR PEABODY ENGINEERING LIMESTONE SCRUBBING STARTUP 1/76	A FOURTH HOT FLUE GAS RUN OF 547 HRS. DURATION WAS TERMINATED ON DECEMBER 29 DUE TO EXCESSIVE VIBRATION OF THE ID BOOSTER FAN. DURING THE COURSE OF THE RUN MINOR OUTAGES OCCURRED PRIOR TO TERMINATION. THE OUTAGES WERE CAUSED BY BOILER SHUTDOWN DUE TO GENERATOR OVERHAUL, INTERRUPTION OF THE FUEL OIL SUPPLY TO THE REHEATER, AND REPAIR OF THE PACKING OF ONE OF THE SCRUBBER PUMPS. THE SYSTEM OPERATED AT 89 PERCENT OF THE DESIGN CAPACITY. SO2 REMOVAL REMAINED AT THE 90 PERCENT LEVEL FOR 1-3 PERCENT SULFUR COAL. THE SYSTEM IS SCHEDULED FOR RESTART IN MID-JANUARY.
I.D. NUMBER 27 DUQUESNE LIGHT ELRAMA 510 MW - RETROFIT COAL 1.0 - 2.8 PERCENT SULFUR CHEMICO LIME SCRUBBING STARTUP 10/75	EQUIPMENT INSTALLATIONS AND MODIFICATIONS WERE COMPLETED. THE SYSTEM STARTED UP IN THE EARLY PART OF OCTOBER 1975. THE INITIAL SHAKEDOWN AND DEBUGGING PHASE PROCEEDED WITH FLUE GAS FROM ONE 100 MW BOILER BEING TREATED BY ONE SCRUBBER. THE UTILITY HAS CHECKED OUT 4 OF THE 5 SCRUBBER MODULES TO DATE. THE FIFTH MODULE WAS INOPERATIVE DUE TO MAINTANENCE TO THE RUBBER-LINED RECYCLE PUMPS. THE IUCS METHOD OF SLUDGE FIXATION IS USED AT THIS STATION.
I.D. NUMBER 28 DUQUESNE LIGHT PHILLIPS 410 MW - RETROFIT COAL 1.0- 2.8 PERCENT SULFUR CHEMICO LIME SCRUBBING STARTUP 7/73	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE JULY, 1973.
I.D. NUMBER 29 GENERAL MOTORS CHEVROLET PARMA 1 2 3 & 4 32 MW - RETROFIT COAL 2.5 PERCENT SULFUR KOCH DOUBLE ALKALI STARTUP 3/74	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE MARCH 1974. SYSTEM AVAILABILITY HAS AVERAGED OVER 68 PERCENT DURING THE PAST 4 MONTHS. SYSTEM AVAILABILITY WAS IN EXCESS OF 58 PERCENT DURING THE REPORT PERIOD.
I.D. NUMBER 30 GULF POWER CO. SCHOLZ NO. 1A 20 MW - RETROFIT COAL 3.0 PERCENT SULFUR ADL/COMBUSTION EQUIP ASSOCIATE DOUBLE ALKALI STARTUP 2/75	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE FEBRUARY,1975. AVAILABILITY HAS AVERAGED OVER 83 PERCENT DURING THE PAST 4 MONTHS. AVAILABILITY FOR THIS SYSTEM WAS 90 PERCENT IN NOVEMBER AND 100 PERCENT IN DECEMBER 1975.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 31 GULF POWER CO. SCHOLZ NO. 2A 20 MW - RETROFIT COAL 3.0 PERCENT SULFUR FOSTER WHEELER ACTIVATED CARBON STARTUP 2/76	THE ADSORPTION AND REGENERATION SECTIONS INITIAL SHAKEDOWN RUN ON FLUE GAS BEGAN ON AUGUST 11 AND WAS CONTINUED FOR A PERIOD OF TEN DAYS. FLOW OF FLUE GAS THROUGH THE ADSORBER WAS CONTINUOUS DURING THAT PERIOD. PRESSURE DROP ACROSS THE ADSORBER WAS WELL BELOW DESIGN LEVEL AND SO2 REMOVAL WAS HIGH. THE REGENERATION SECTION WAS OPERATIONAL ABOUT 60 PERCENT DURING THE PERIOD. THE RESOX PORTION OF THE SYSTEM RAN FOR 5 DAYS IN THE LATTER PART OF OCTOBER. THE RESOX RUN WAS SUCCESSFUL. THE UTILITY IS NOW IN THE PROCESS OF MAKING MODIFICATIONS FOR A SCHEDULED FEBRUARY STARTUP.
I.D. NUMBER 32 GULF POWER CO. SCHOLZ NOS. 1B & 2B 23 MW - RETROFIT COAL 3.0 PERCENT SULFUR CHIYODA INTERNATIONAL THOROUGHBRED 101 STARTUP 3/75	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE MARCH,1975. AVAILABILITY HAS AVERAGED 85 PERCENT FOR THE PAST 4 MONTHS. AVAILABILITY WAS ALMOST 100 PERCENT IN NOVEMBER AND 98 PERCENT IN DECEMBER. SO2 REMOVAL EFFICIENCY IS REPORTED TO BE IN EXCESS OF 95 PERCENT.
I.D. NUMBER 33 INDIANAPOLIS POWER & LIGHT CO. PETERSBURG NO 3 530 MW - NEW COAL 3.0-3.5 PERCENT SULFUR UNIVERSAL OIL PRODUCTS LIMESTONE SCRUBBING STARTUP 4/77	UNIVERSAL OIL PRODUCTS HAS BEEN SELECTED TO BUILD THE LIMESTONE SCRUBBING SYSTEM. THE ENTIRE 530MW UNIT WILL BE TREATED BY FOUR MODULES. CONSTRUCTION HAS BEGUN ON ALL FOUR MODULES. WORK ON THE FOUNDATION IS PROCEEDING.
I.D. NUMBER 34 KANSAS CITY POWER & LIGHT HAWTHORN NO 3 140 MW - RETROFIT COAL 0.6- 3.0 PERCENT SULFUR COMBUSTION ENGINEERING LIMESTONE INJECTION &WET SCRUB STARTUP 11/72	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE NOVEMBER,1972. AVAILABILITY FOR BOTH MODULES WAS ZERO DURING THE PAST TWO REPORTING MONTHS. THE BOILER AND SCRUBBING MODULES WERE SHUTDOWN DURING THE MONTHS OF NOVEMBER AND DECEMBER FOR A SCHEDULED TURBINE OVERHAUL. THE SYSTEM IS SCHEDULED TO GO BACK ON-LINE AGAIN SOMETIME IN JANUARY 1976. SYSTEM AVAILABILITY FOR MODULES A AND B DURING THE PERIOD OF SEPTEMBER-OCTOBER 1975 WAS 67 AND 52 PERCENT RESPECTIVELY.
I.D. NUMBER 35 KANSAS CITY POWER & LIGHT HAWTHORN NO 4 100 MW - RETROFIT COAL 0.6- 3.0 PERCENT SULFUR COMBUSTION ENGINEERING LIMESTONE INJECTION &WET SCRUB STARTUP 8/72	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. FGD SYSTEM HAS BEEN IN OPERATION SINCE AUGUST,1972. AVAILABILITY FOR MODULES A AND B HAS AVERAGED 0 AND 58 PERCENT RESPECTIVELY FOR THE PAST TWO MONTHS. MODULE A WAS UNAVAILABLE TROUGHOUT THE REPORT PERIOD. AVAILABILITY FOR MODULE B WAS 99 PERCENT IN NOVEMBER AND 16 PERCENT IN DECEMBER. AVERAGE AVAILABILITY FOR MODULES A AND B DURING THE PAST 4 MONTHS HAS BEEN 28 AND 65 PERCENT RESPECTIVELY.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 36 KANSAS CITY POWER & LIGHT LA CYGNE NO 1 820 MW - NEW COAL 5.2 PERCENT SULFUR BABCOCK & WILCOX LIMESTONE SCRUBBING STARTUP 2/73	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE DECEMBER,1968. AVERAGE AVAILABILITY FOR ALL 7 MODULES HAS BEEN 84 PERCENT FOR THE PAST 8 MONTHS. AVERAGE SYSTEM AVAILABILITY WAS 91 PERCENT IN NOVEMBER AND 86 PERCENT IN DECEMBER.
I.D. NUMBER 37 KANSAS POWER & LIGHT JEFFERY NO. 1 700 MW - NEW COAL COMBUSTION ENGINEERING LIMESTONE SCRUBBING STARTUP 6/78	WORK ON THE FGD SYSTEM IS IN THE EARLY PLANNING STAGE. COMBUSTION ENGINEERING IS LOOKING INTO THE SO2 EMISSION CONTROL NEEDS OF THIS PLANT AND IS PREPARING A PROPOSAL TO KPL. CONTROL OF PARTICULATES WILL BE ACCOMPLISHED IN TWO ESP UNITS TO BE INSTALLED UPSTREAM OF THE EIGHT FGD MODULES.
I.D. NUMBER 38 KANSAS POWER & LIGHT JEFFERY NO. 2 700 MW - NEW COAL COMBUSTION ENGINEERING LIMESTONE SCRUBBING STARTUP 6/79	WORK ON THE FGD SYSTEM IS IN THE EARLY PLANNING STAGE. COMBUSTION ENGINEERING IS LOOKING INTO THE SO2 EMISSION CONTROL NEEDS OF THIS PLANT AND IS PREPARING A PROPOSAL TO KPL. CONTROL OF PARTICULATES WILL BE ACCOMPLISHED IN TWO ESP UNITS TO BE INSTALLED UPSTREAM OF EIGHT FGD MODULES.
I.D. NUMBER 39 KANSAS POWER & LIGHT LAWRENCE NO 4 125 MW - RETROFIT COAL 3.5 PERCENT SULFUR COMBUSTION ENGINEERING LIMESTONE INJECTION &WET SCRUB STARTUP 12/68	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE DECEMBER,1968.
I.D. NUMBER 40 KANSAS POWER & LIGHT LAWRENCE NO 5 400 MW - NEW COAL 0.5 PERCENT SULFUR COMBUSTION ENGINEERING LIMESTONE INJECTION &WET SCRUB STARTUP 11/71	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE NOVEMBER,1971.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 41 KENTUCKY UTILITIES GREEN RIVER UNITS 1 AND 2 64 MW - RETROFIT COAL 3.8 PERCENT SULFUR AMERICAN AIR FILTER LIME SCRUBBING STARTUP 9/75	THE FGD SYSTEM AT THIS FACILITY IS DESIGNED TO TREAT THE FLUE GAS GENERATED FROM THREE,HIGH SULFUR,COAL-FIRED BOILERS WHICH PRODUCE A TOTAL FLUE GAS CAPACITY EQUIVALENT TO 64 MW. THE FLUE GAS IS FED TO THE SCRUBBING SYSTEM VIA A COMMON HEADER. THE SYSTEM BECAME OPERATIONAL SEPTEMBER 13,1975. SYSTEM OPERATION IS CURRENTLY BEING CONDUCTED ON A HALF-LOAD BASIS BECAUSE OF A TURBINE OVERHAUL. ADDITIONAL INFORMATION ON THIS SYSTEM IS AVAILABLE IN THE BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT.
I.D. NUMBER 42 KEY WEST UTILITY BOARD STOCK ISLAND PLANT 37 MW - NEW OIL 2.4 PERCENT SULFUR ZURN AIR SYSTEMS LIMESTONE SCRUBBING STARTUP 10/72	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE OCTOBER,1972. FGD SYSTEM AVAILABILITY HAS BEEN 0 PERCENT SINCE JANUARY,1975.
I.D. NUMBER 43 LOUISVILLE GAS & ELECTRIC CANE RUN NO 1 110 MW - RETROFIT COAL 3.5- 4.0 PERCENT SULFUR NOT SELECTED LIME SCRUBBING STARTUP 6/80	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT ESTABLISHING 6/80 AS THE STARTUP DATE FOR A LIME SCRUBBING FGD SYSTEM.
I.D. NUMBER 44 LOUISVILLE GAS & ELECTRIC CANE RUN NO 2 107 MW - RETROFIT COAL 3.5- 4.0 PERCENT SULFUR NOT SELECTED LIME SCRUBBING STARTUP 6/80	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
I.D. NUMBER 45 LOUISVILLE GAS & ELECTRIC CANE RUN NO 3 137 MW - RETROFIT COAL 3.5- 4.0 PERCENT SULFUR NOT SELECTED LIME SCRUBBING STARTUP 6/80	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/80 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 46 LOUISVILLE GAS & ELECTRIC CANE RUN NO 4 178 MW - RETROFIT COAL 3.5-4.05 PERCENT SULFUR AMERICAN AIR FILTER LIME SCRUBBING STARTUP 6/76	THE UTILITY HAS SIGNED A CONSENT AGREEMENT WITH THE EPA FOR THE INSTALLATION AND OPERATION OF FGD EQUIPMENT ON THIS UNIT BY JUNE 1,1976. THE CONTRACT HAS BEEN AWARDED TO AMERICAN AIR FILTER. THE SYSTEM WILL OPERATE ON A CLOSED WATER LOOP AND THE SLUDGE WILL BE STABILIZED. THIS SYSTEM IS PRESENTLY UNDER CONSTRUCTION.
I.D. NUMBER 47 LOUISVILLE GAS & ELECTRIC CANE RUN NO 5 183 MW - RETROFIT COAL 3.5-4.05 PERCENT SULFUR COMBUSTION ENGINEERING LIME SCRUBBING STARTUP 12/77	THE UTILITY HAS SIGNED A CONSENT AGREEMENT WITH THE EPA FOR THE INSTALLATION AND OPERATION OF FGD EQUIPMENT ON THIS UNIT BY DECEMBER 1,1977. THE CONTRACT FOR THIS LIME SCRUBBING SYSTEM HAS BEEN AWARDED TO COMBUSTION ENGINEERING. THE UNIT IS PRESENTLY UNDER CONSTRUCTION.
I.D. NUMBER 48 LOUISVILLE GAS & ELECTRIC CANE RUN NO 6 277 MW - RETROFIT COAL 3.5- 4.0 PERCENT SULFUR NOT SELECTED LIME SCRUBBING STARTUP 9/78	THE UTILITY HAS SIGNED A CONSENT AGREEMENT WITH THE EPA FOR THE INSTALLATION AND OPERATION OF FGD EQUIPMENT ON THIS UNIT BY SEPTEMBER 1,1978. THE UNIT IS PRESENTLY OUT FOR BIDS.
I.D. NUMBER 49 LOUISVILLE GAS & ELECTRIC MILL CREEK NO 1 330 MW - RETROFIT COAL 3.5- 4.0 PERCENT SULFUR NOT SELECTED LIME SCRUBBING STARTUP 6/81	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/81 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.
I.D. NUMBER 50 LOUISVILLE GAS & ELECTRIC MILL CREEK NO 2 330 MW - RETROFIT COAL 3.5- 4.0 PERCENT SULFUR NOT SELECTED LIME SCRUBBING STARTUP 6/82	A COMPLIANCE SCHEDULE HAS BEEN SUBMITTED TO THE JEFFERSON COUNTY AIR POLLUTION CONTROL DISTRICT WITH 6/82 ESTABLISHED AS THE STARTUP DATE FOR AN FGD SYSTEM.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 51 LOUISVILLE GAS & ELECTRIC MILL CREEK NO 3 425 MW - NEW COAL 3.5- 4.0 PERCENT SULFUR AMERICAN AIR FILTER LIME SCRUBBING STARTUP 7/77	THE UTILITY HAS SIGNED A CONSENT AGREEMENT WITH THE EPA FOR THE INSTALLATION AND OPERATION OF FGD EQUIPMENT ON THIS POWER GENERATING UNIT BY JULY 1,1977. THIS NEW FACILITY WILL BE REQUIRED TO MEET FEDERAL NSPS FOR THE LIMITATION OF SO2 EMISSIONS. CONTRACT HAS BEEN AWARDED TO AMERICAN AIR FILTER. THE SYSTEM WILL OPERATE ON A CLOSED WATER LOOP AND THE SLUDGE WILL BE STABILIZED. CONSTRUCTION HAS BEGUN ON THE FGD SYSTEM.
I.D. NUMBER 52 LOUISVILLE GAS & ELECTRIC MILL CREEK NO 4 425 MW - NEW COAL 3.5- 4.0 PERCENT SULFUR NOT SELECTED LIME SCRUBBING STARTUP 7/79	THE UTILITY HAS SIGNED A CONSENT AGREEMENT WITH THE EPA FOR THE INSTALLATION AND OPERATION OF FGD EQUIPMENT ON THIS POWER GENERATING UNIT BY JULY 1,1979. THIS NEW FACILITY WILL BE REQUIRED TO MEET FEDERAL NSPS FOR THE LIMITATION OF SO2 EMISSIONS. THE UNIT IS PRESENTLY OUT FOR BIDS.
I.D. NUMBER 53 LOUISVILLE GAS & ELECTRIC PADDYS RUN NO 6 65 MW - RETROFIT COAL 3.5-4.0 PERCENT SULFUR COMBUSTION ENGINEERING LIME SCRUBBING STARTUP 4/73	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE APRIL 1973. AVAILABILITY WAS 100 PERCENT IN NOVEMBER AND 90 PERCENT IN DECEMBER. AVERAGE AVAILABILITY FOR THE PAST 4 MONTHS HAS EXCEEDED 97 PERCENT. SO2 REMOVAL EFFICIENCY IS REPORTED TO BE IN EXCESS OF 98 PERCENT.
I.D. NUMBER 54 MINNKOTA POWER COOPERATIVE MILTON R. YOUNG NO. 2 450 MW - NEW LIGNITE - 6,500 BTU, 0.7% S COMBUSTION EQUIP. ASSOCIATES LIME SCRUBBING STARTUP 0/77	BOILER IS UNDER CONSTRUCTION. FGD CONRACT HAS BEEN AWARDED AND ENGINEERING WORK BY SANDERSON AND PORTER IS IN PROGRESS. A SMALL PILOT PLANT TREATING 5,000 ACFM WENT INTO OPERATION ON OCTOBER 13,1975 AND WILL RUN INTO DECEMBER. NO RESULTS ARE AVAILABLE.
I.D. NUMBER 55 MONTANA POWER CO. COLSTRIP NO 1 360 MW - NEW COAL 0.8 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES LIME SCRUBBING STARTUP 10/75	CONSTRUCTION WAS COMPLETED ON SCHEDULE. THE SYSTEM STARTED UP IN OCTOBER AND PROCEEDED THROUGH A DEBUGGING AND SHAKEDOWN PHASE. THIS UNIT HAS GONE FULL COMMERCIAL OPERATION AND IS PRESENTLY TREATING THE EQUIVALENT OF 230 MW OF FLUE GAS (TOTAL DESIGN CAPACITY IS 360 MW).

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 56 MONTANA POWER CO. COLSTRIP NO 2 360 MW - NEW COAL 0.8 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES LIME SCRUBBING STARTUP 7/76	FGU SYSTEM UNDER CONSTRUCTION WITH NO MAJOR DELAYS. ANTICIPATE COMPLETING CONSTRUCTION ON SCHEDULE. TURBINE ROLL AND SYSTEM CHECKOUT SCHEDULED TO COMMENCE IN MAY 1976. NO DELAYS OF SCHEDULED STARTUP DATE FORESEEN AT THIS TIME.
I.D. NUMBER 57 MONTANA POWER CO. COLSTRIP NO.3 700 MW - NEW COAL 0.7 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 7/79	UTILITY IS WAITING FOR APPROVAL OF A CERTIFICATE TO CONSTRUCT BOILERS NO.3 AND NO.4 BEFORE PROCEEDING WITH PLANS TO INSTALL FGD SYSTEMS.
I.D. NUMBER 58 MONTANA POWER CO. COLSTRIP NO.4 700 MW - NEW COAL 0.7 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 7/80	UTILITY IS WAITING FOR APPROVAL OF A CERTIFICATE TO CONSTRUCT BOILERS NO.3 AND NO.4 BERFORE PROCEEDING WITH PLANS TO INSTAL FGD SYSTEMS.
I.D. NUMBER 59 NEVADA POWER HARRY ALLEN STATION NO. 1 500 MW - NEW COAL NOT SELECTED NOT SELECTED STARTUP 6/82	CONSIDERING HOT SIDE ESP IN CONJUNCTION WITH AN FGD SYSTEM. SPECIFICATIONS HAVE NOT YET BEEN PREPARED.
I.D. NUMBER 60 NEVADA POWER HARRY ALLEN STATION NO. 2 500 MW - NEW COAL NOT SELECTED NOT SELECTED STARTUP 6/83	CONSIDERING HOT SIDE ESP IN CONJUNCTION WITH AN FGD SYSTEM. SPECIFICATIONS HAVE NOT YET BEEN PREPARED.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 61 NEVADA POWER HARRY ALLEN STATION NO. 3 500 MW - NEW COAL NOT SELECTED NOT SELECTED STARTUP 6/84	CONSIDERING HOT SIDE ESP IN CONJUNCTION WITH AN FGD SYSTEM. SPECIFICATIONS HAVE NOT YET BEEN PREPARED.
I.D. NUMBER 62 NEVADA POWER HARRY ALLEN STATION NO. 500 MW - NEW COAL NOT SELECTED NOT SELECTED STARTUP 6/85	CONSIDERING HOT SIDE ESP IN CONJUNCTION WITH AN FGD SYSTEM. SPECIFICATIONS HAVE NOT YET BEEN PREPARED.
I.D. NUMBER 63 NEVADA POWER REID GARDNER NO 1 125 MW - RETROFIT COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 4/74	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE APRIL 1974.
I.D. NUMBER 64 NEVADA POWER REID GARDNER NO 2 125 MW - RETROFIT COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 12/73	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS FGD SYSTEM HAS BEEN OPERATIONAL SINCE DECEMBER 1973.
I.D. NUMBER 65 NEVADA POWER REID GARDNER NO 3 125 MW - NEW COAL 0.5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 6/76	REFER TO MONTHLY COMMENTS IN BACKGROUND INFORMATION SECTION INCLUDED WITH TABLE 3 FOR THIS UNIT.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 66 NEVADA POWER REID GARDNER NO 4 125 MW - NEW COAL 0-5- 1.0 PERCENT SULFUR COMBUSTION EQUIP. ASSOCIATES SODIUM CARBONATE SCRUBBING STARTUP 0/ 0	NEVADA POWER COMPANY HAS SIGNED A LETTER OF INTENT WITH COMBUSTION EQUIPMENT ASSOCIATES FOR THE CONSTRUCTION OF AN FGD SYSTEM ON REID GARDNER NO. 4. HOWEVER, CONSTRUCTION OF THE BOILER HAS BEEN INDEFINITELY POSTPONED.
I.D. NUMBER 67 NEVADA POWER WARNER VALLEY STATION NO. 1 250 MW - NEW COAL NOT SELECTED NOT SELECTED STARTUP 6/81	SPECIFICATIONS ARE BEING PREPARED FOR A SCRUBBING SYSTEM.
I.D. NUMBER 68 NEVADA POWER WARNER VALLEY STATION NO. 2 250 MW - NEW COAL NOT SELECTED NOT SELECTED STARTUP 6/82	SPECIFICATIONS ARE BEING PREPARED FOR A SCRUBBING SYSTEM.
I.D. NUMBER 69 NEW ENGLAND ELEC SYSTEM BRAYTON POINT NO.3 650 MW - RETROFIT COAL 3.0 PERCENT SULFUR NOT SELECTED REGENERABLE NOT SELECTED STARTUP 0/ 0	INVESTIGATING THE POSSIBILITY OF AN ADVANCED REGENERABLE FGD SYSTEM HAVING ELEMENTAL SULFUR AS AN END PRODUCT WHICH HAS PROMISE OF A BREAKTHROUGH IN CAPITAL AND OPERATING COST.
I.D. NUMBER 70 NORTHERN INDIANA PUB SERVICE BAILLY NO. 7 190 MW - RETROFIT COAL 3 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 0/ 0	CONSIDERING A LIME OR LIMESTONE SCRUBBING UNIT. ALSO WAITING FOR PERFORMANCE OF WELLMAN LORD/ALLIED CHEMICAL UNIT UNDER CONSTRUCTION AT THEIR D. H. MITCHEEL NO. 11 UNIT. LOW SULFUR COAL MAY BE BURNED TO COMPLY WITH SO2 EMISSION REGULATIONS.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 71 NORTHERN INDIANA PUB SERVICE BAILLY NO. 8 400 MW - RETROFIT COAL 3 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 0/ 0	CONSIDERING A LIME OR LIMESTONE SCHUBBING UNIT. ALSO WAITING FOR PERFORMANCE OF WELLMAN LORD/ALLIED CHEMICAL UNIT UNDER CONSTRUCTION AT THEIR D. H. MITCHELL NO. 11 UNIT. LOW SULFUR COAL MAY BE BURNED TO COMPLY WITH SO2 EMISSION REGULATIONS.
I.D. NUMBER 72 NORTHERN INDIANA PUB SERVICE MITCHELL NO 11 115 MW - RETROFIT COAL 3.2- 3.5 PERCENT SULFUR DAVY POWERGAS/ALLIED CHEMICAL WELLMAN LORD/ALLIED CHEMICAL STARTUP 3/76	SULFUR DIOXIDE WILL BE CONVERTED TO ELEMENTAL SULFUR IN AN ADJACENT PLANT USING THE ALLIED CHEMICAL SO2 REDUCTION PROCESS. THIS INSTALLATION IS PARTIALLY FUNDED BY AN EPA DEMONSTRATION GRANT. THIS UNIT IS STILL UNDER CONSTRUCTION. THE STARTUP DATE HAS BEEN MOVED BACK TO MARCH 1976.
I.D. NUMBER 73 NORTHERN STATES POWER CO. SHERBURNE NO 1 680 MW - NEW COAL 0.8 PERCENT SULFUR COMBUSTION ENGINEERING LIMESTONE SCRUBBING STARTUP 5/76	FGD SYSTEM IS STILL UNDER CONSTRUCTION, PURCHASE ORDERS FOR SCRUBBERS WERE PLACED IN EARLY 1971. DESIGN MODIFICATIONS CAUSED EQUIPMENT DELIVERY DELAYS,HOWEVER THE OVERALL SCHEDULE HAS NOT BEEN UPSET. PROBLEMS ENCOUNTERED BY THE UTILITY THUS FAR HAVE BEEN DUE TO OBTAINING SOME ELECTRICAL CONTROLS, A RECENT CONSTRUCTION STRIKE, AND A BOILER MODIFICATION INVOLVING GREATER NEGATIVE PRESSURE TOLERANCE. THE UTILITY IS NOW BEGINNING A PRELIMINARY SYSTEM CHECKOUT BY PASSING AIR AND CLEAR WATER THROUGH THE FGD UNIT. A MAY STARTUP OF THIS SYTEM IS STILL ANTICIPATED.
I.D. NUMBER 74 NORTHERN STATES POWER CO. SHERBURNE NO 2 680 MW - NEW COAL 0.8 PERCENT SULFUR COMBUSTION ENGINEERING LIMESTONE SCRUBBING STARTUP 5/77	FGD SYSTEM UNDER CONSTRUCTION WITH NO MAJOR DELAYS. PURCHASE ORDERS FOR SCRUBBERS WERE PLACED IN EARLY 1971. RECENT SCRUBBER DESIGN CHANGES HAVE CAUSED EQUIPMENT DELIVERY DELAYS BUT THE OVERALL SCHEDULE HAS NOT BEEN UPSET. PROBLEMS HAVE ARISEN IN OBTAINING SOME ELECTRICAL CONTROLS. CONSTRUCTION IS BEHIND SCHEDULE AND SOME DELAY OF STARTUP DATE MAY BE EXPECTED DUE TO RECENT STRIKE. THE EXTENT OF THE DELAY CANNOT BE DETERMINED AT THE PRESENT TIME.
I.D. NUMBER 75 PENNSYLVANIA POWER CO. BRUCE MANSFIELD NO. 1 835 MW - NEW COAL 4.3 PERCENT SULFUR CHEMICO LIME SCRUBBING STARTUP 4/76	THE UTILITY BEGAN A SHAKEDOWN AND DEBUGGING PHASE OF OPERATION FOR PART OF THE SYSTEM IN DECEMBER 1975. CONSTRUCTION IS STILL IN PROGRESS ON THE REMAINING MODULES OF THIS SYSTEM. FULL COMMERCIAL STARTUP IS PRESENTLY PROJECTED FOR APRIL 1976.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 76 PENNSYLVANIA POWER CO. BRUCE MANSFIELD NO. 2 835 MW - NEW COAL 4.3 PERCENT SULFUR CHEMICO LIME SCRUBBING STARTUP 4/77	FGD SYSTEM UNDER CONSTRUCTION, BUT ENCOUNTERING LONG DELAYS IN MATERIAL DELIVERIES. STARTUP DATE HAS BEEN DELAYED 1 YEAR UNTIL 4/77.
I.D. NUMBER 77 PENNSYLVANIA POWER CO. BRUCE MANSFIELD NO. 3 835 MW - NEW COAL 4.3 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 4/79	FGD UNIT IS PLANNED. PRESENTLY REQUESTING AND EVALUATING BIDS.
I.D. NUMBER 78 PHILADELPHIA ELECTRIC CO. CROMBY 150 MW - RETROFIT COAL 2-4 PERCENT SULFUR UNITED ENGINEERS MAGNESIUM OXIDE SCRUBBING STARTUP 10/78	COMPANY PLAN TO RETROFIT ONE OF THE TWO BOILERS AT CROMBY BUT HAVE NOT MADE A FINAL DECISION ON THIS POINT. THE FGD PROCESS SELECTED WILL VERY LIKELY BE AN MGO PROCESS DESIGNED JOINTLY BY UNITED ENGINEERS AND PHILADELPHIA ELECTRIC. ENGINEERING DESIGN WORK IS SCHEDULED FOR JAN 76. STARTUP NOW SCHEDULED FOR OCT 78.
I.D. NUMBER 79 PHILADELPHIA ELECTRIC CO. EDDYSTONE NO 1A 120 MW - RETROFIT COAL 2.5 PERCENT SULFUR UNITED ENGINEERS MAGNESIUM OXIDE SCRUBBING STARTUP 9/75	THE SO2 SCRUBBER HAS BEEN TEMPORARILY SHUTDOWN BECAUSE THE ACID PLANT REGENERATION FACILITY AT THE OLIN CHEMICAL SULFURIC ACID PLANT IN PAULSBORO NEW JERSEY HASPERMANENTLYCEASED OPERATIONS. CONSIDERATIONS ARE NOW BEING GIVEN TO THE RELOCATION OF THE REGENERATION FACILITY. ONCE THIS DECISION IS MADE A MINIMUM PERIOD OF SIX MONTHS WILL BE REQUIRED TO RELOCATE THE REGENERATION FACILITY. THE PARTICULATE SCRUBBERS ARE CONTINUING TO OPERATE. THE UTILITY IS STILL EXPERIENCING PROBLEMS WITH THE FANS, REHEAT BURNERS, DAMPERS, AND EXPANSION JOINTS.
I.D. NUMBER 80 PHILADELPHIA ELECTRIC CO. EDDYSTONE NO 1B 240 MW - RETROFIT COAL 2.5 PERCENT SULFUR UNITED ENGINEERS MAGNESIUM OXIDE SCRUBBING STARTUP 10/78	THE RETROFITTING OF SECOND STAGE SCRUBBERS ON THE BALANCE OF THE FLUE GAS FROM EDDYSTONE 1 IS AWAITING PERFORMANCE OF THE EXISTING 2-STAGE UNIT WHICH IS TREATING ONE THIRD OF THE BOILER FLUE GAS. THIS UNIT CONSISTS OF A AND B PARTICULATE SCRUBBING TRAINS, EACH ARE 120 MW. CHECK VALVE FAILURE ON A TRAIN HAS BEEN CORRECTED. FAN ON B TRAIN IS STILL AT THE FACTORY FOR REPAIR.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 81 PHILADELPHIA ELECTRIC CO. EDDYSTONE NO 2 336 MW - RETROFIT COAL 2.4 PERCENT SULFUR UNITED ENGINEERS MAGNESIUM OXIDE SCRUBBING STARTUP 10/78	COMPANY IS AWAITING PERFORMANCE OF FGD SYSTEM ON UNIT 1 BEFORE PROCEEDING WITH DESIGN OF FGD SYSTEM ON THIS BOILER. HOWEVER ENGINEERING PHASE IS TENTATIVELY PLANNED TO START IN JAN 76. AT WHICH TIME EVALUATION OF DATA FROM UNIT 1 SHOULD BE COMPLETE.
I.D. NUMBER 82 PUBLIC SERVICE CO OF NEW MEX. SAN JUAN NO. 1 375 MW - NEW COAL 0.8 PERCENT SULFUR DAVY POWERGAS/ALLIED CHEMICAL WELLMAN LORD/ALLIED CHEMICAL STARTUP 7/77	FGD PROJECT IS PROCEEDING ACCORDING TO PLAN WITH NO MAJOR PROBLEMS. EQUIPMENT IS PRESENTLY BEING PROCURED. CONSTRUCTION TO BEGIN IN MARCH OR APRIL 1976.
I.D. NUMBER 83 PUBLIC SERVICE CO OF NEW MEX. SAN JUAN NO. 2 340 MW - RETROFIT COAL 0.8 PERCENT SULFUR DAVY POWERGAS/ALLIED CHEMICAL WELLMAN LORD/ALLIED CHEMICAL STARTUP 7/77	FGD PROJECT IS PROCEEDING ACCORDING TO PLAN WITH NO MAJOR PROBLEMS. EQUIPMENT IS PRESENTLY BEING PROCURED. CONSTRUCTION TO BEGIN IN MARCH OR APRIL 1976.
I.D. NUMBER 84 PUBLIC SERVICE CO OF NEW MEX. SAN JUAN NO. 3 500 MW - NEW COAL 0.8 PERCENT SULFUR DAVY POWERGAS WELLMAN LORD STARTUP 5/78	A LETTER OF INTENT HAS BEEN SIGNED WITH DAVY POWERGAS FOR A WELLMAN LORD UNIT.
I.D. NUMBER 85 PUBLIC SERVICE CO OF NEW MEX. SAN JUAN NO. 4 500 MW - NEW COAL 0.8 PERCENT SULFUR DAVY POWERGAS WELLMAN LORD STARTUP 5/80	A LETTER OF INTENT HAS BEEN SIGNED WITH DAVY POWERGAS FOR A WELLMAN LORD UNIT.

```
 COMPANY                                        TABLE 2
                                 STATUS OF FGD SYSTEMS DURING  12/75

 POWER STATION                                                CURRENT MONTH
-------------------------------------------------------------------------------------------------------------------------
I.D. NUMBER   86                      REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS
PUBLIC SERVICE CO. OF COLORADO        SYSTEM HAS BEEN OPERATIONAL SINCE NOVEMBER 1974.
VALMONT NO. 5
   50  MW - RETROFIT
COAL  0.72 PERCENT SULFUR
UOP / PUB SERVICE OF COLORADO
LIMESTONE SCRUBBING
STARTUP  10/74
-------------------------------------------------------------------------------------------------------------------------
I.D. NUMBER   87
PUBLIC SERVICE INDIANA                BID EVALUATION IS UNDERWAY.
GIBSON NO. 3
  650  MW - NEW
COAL, 3.3% SULFUR
NOT SELECTED
NOT SELECTED
STARTUP   0/78
-------------------------------------------------------------------------------------------------------------------------
I.D. NUMBER   88
PUBLIC SERVICE INDIANA                BID SPECIFICATIONS ARE BEING PREPARED.
GIBSON NO. 4
  650  MW - NEW
COAL
NOT SELECTED
NOT SELECTED
STARTUP   0/79
-------------------------------------------------------------------------------------------------------------------------
I.D. NUMBER   89                      CONSTRUCTION OF THIS SYSTEM IS PRESENTLY PROCEEDING ON SCHEDULE. WORK ON
RICKENBACKER AFB                      THE SLUDGE POND, FOUNDATION, AND TOWER IS NEARING COMPLETION. THE
RICKENBACKER                          SCHEDULED FEBRUARY STARTUP DATE IS STILL OPERATIVE.
   20  MW - RETROFIT
COAL   5.0 PERCENT SULFUR
RESEARCH COTTRELL
LIME SCRUBBING
STARTUP   2/76
-------------------------------------------------------------------------------------------------------------------------
I.D. NUMBER   90                      A LETTER OF INTENT HAS BEEN SIGNED WITH B&W FOR CONSTRUCTION OF A LIME-
S. CAROLINA PUB SERV AUTHORITY        STONE SYSTEM TREATING 140MW OF THE TOTAL 280MW.  DIFFERENT METHODS OF
WINYAH NO. 2                          DISPOSAL ARE BEING EXAMINED.
  140  MW - NEW
COAL  1.0 PERCENT SULFUR
BABCOCK & WILCOX
LIMESTONE SCRUBBING
STARTUP   5/77
-------------------------------------------------------------------------------------------------------------------------
```

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 91 S. MISSISSIPPI ELEC PWR ASSOC R.D. MORROW NO.1 180 MW - NEW COAL 1 PERCENT SULFUR RILEY STOKER/ENVIRONEERING LIMESTONE SCRUBBING STARTUP 11/77	THE CONTRACT FOR THIS LIMESTONE SYSTEM HAS BEEN AWARDED TO RILEY STOKER/ENVIRONEERING.
I.D. NUMBER 92 S. MISSISSIPPI ELEC PWR ASSOC R.D. MORROW NO.2 180 MW - NEW COAL 1 PERCENT SULFUR RILEY STOKER/ENVIRONEERING LIMESTONE SCRUBBING STARTUP 6/78	THE CONTRACT FOR THIS LIMESTONE SYSTEM HAS BEEN AWARDED TO RILEY STOKER/ENVIRONEERING.
I.D. NUMBER 93 SOUTHERN CALIFORNIA EDISON KAIPAROWITZ NO. 1 750 MW - NEW COAL 10,800,BTU, 0.5% S NOT SELECTED LIME SCRUBBING STARTUP 0/82	THIS NEW STATION IS ANTICIPATED TO INCORPORATE FGD, BUT A SPECIFIC PROCESS HAS NOT BEEN SELECTED. THE STARTUP DATE FOR THIS UNIT HAS BEEN DELAYED ONE YEAR PENDING THE ASSEMBLY AND EVALUATION OF DATA FOR ENVIRONMENTAL CONSIDERATIONS AND LENGTHY APPROVAL PROCESSES.
I.D. NUMBER 94 SOUTHERN CALIFORNIA EDISON KAIPAROWITZ NO. 2 750 MW - NEW COAL 10,800,BTU, 0.5% S NOT SELECTED LIME SCRUBBING STARTUP 0/83	THIS NEW STATION IS ANTICIPATED TO INCORPORATE FGD, BUT A SPECIFIC PROCESS HAS NOT BEEN SELECTED. THE STARTUP DATE FOR THIS UNIT HAS BEEN DELAYED ONE YEAR PENDING THE ASSEMBLY AND EVALUATION OF DATA FOR ENVIRONMENTAL CONSIDERATIONS AND LENGTHY APPROVAL PROCESSES.
I.D. NUMBER 95 SOUTHERN CALIFORNIA EDISON KAIPAROWITZ NO. 3 750 MW - NEW COAL 10,800,BTU, 0.5% S NOT SELECTED LIME SCRUBBING STARTUP 0/84	THIS NEW STATION IS ANTICIPATED TO INCORPORATE FGD, BUT A SPECIFIC PROCESS HAS NOT BEEN SELECTED. THE STARTUP DATE FOR THIS UNIT HAS BEEN DELAYED ONE YEAR PENDING THE ASSEMBLY AND EVALUATION OF DATA FOR ENVIRONMENTAL CONSIDERATIONS AND LENGTHY APPROVAL PROCESSES.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 96 SOUTHERN CALIFORNIA EDISON KAIPAROWITZ NO. 4 750 MW - NEW COAL 10,800,BTU, 0.5% S NOT SELECTED LIME SCRUBBING STARTUP 0/85	THIS NEW STATION IS ANTICIPATED TO INCORPORATE FGD, BUT A SPECIFIC PROCESS HAS NOT BEEN SELECTED. THE STARTUP DATE FOR THIS UNIT HAS BEEN DELAYED ONE YEAR PENDING THE ASSEMBLY AND EVALUATION OF DATA FOR ENVIRONMENTAL CONSIDERATIONS AND LENGTHY APPROVAL PROCESSES.
I.D. NUMBER 97 SOUTHERN CALIFORNIA EDISON MOHAVE NO 1B 620 MW - RETROFIT COAL 0.5- 0.8 PERCENT SULFUR NOT SELECTED LIME/LIMESTONE SCRUBBING STARTUP 6/77	THE TOTAL CAPACITY OF BOILER IS 790 MW, EXISTING EXPERIMENTAL FGD MODULE WILL TREAT ONLY 170 MW. FINAL SELECTION OF THE PROCESS HAS BEEN POSTPONED. SEE MOHAVE 1A.
I.D. NUMBER 98 SOUTHERN CALIFORNIA EDISON MOHAVE NO. 2 790 MW - RETROFIT COAL 0.5 TO 0.8 % SULFUR NOT SELECTED LIME/LIMESTONE SCRUBBING STARTUP 6/77	A PROTOTYPE 170 MW LIME SCRUBBING SYSTEM (HORIZONTAL MODULE) ON THIS UNIT WAS OPERATED FROM NOVEMBER 1973 UNTIL FEBRUARY 1975. AVAILABILITY FOR THE PERIOD 1/16/74 TO 2/9/75 AVERAGED 73.5%. THIS SYSTEM IS TO BE REINSTALLED AT THE FOUR CORNERS PLANT, OPERATED BY ARIZONA PUBLIC SERVICE.
I.D. NUMBER 99 SPRINGFIELD CITY UTILITIES SOUTHWEST NO. 1 200 MW - NEW COAL 3.5 PERCENT SULFUR UNIVERSAL OIL PRODUCTS LIMESTONE SCRUBBING STARTUP 6/76	BOTH BOILER AND FGD SYSTEM ARE PRESENTLY UNDER CONSTRUCTION. EXPERIENCING SOME DELAYS BUT COMPLETION IS STILL EXPECTED BY JUNE. SCRUBBER AND ESP ARE NOW BEING ERECTED. ID FAN IS ALSO BEING INSTALLED
I.D. NUMBER 100 TENNESSEE VALLEY AUTHORITY SHAWNEE NO.10A 10 MW - RETROFIT COAL 2.9 PERCENT SULFUR UNIVERSAL OIL PRODUCTS LIME/LIMESTONE SCRUBBING STARTUP 4/72	REFER TO BACKGROUND INFORMATION SECTION IN TABLE 3 OF THIS REPORT. THIS TURBULENT CONTACT ABSORBER(TCA) LIME/LIMESTONE SCRUBBING SYSTEM HAS BEEN OPERATIONAL SINCE APRIL 1972. THIS TEST PROGRAM IS FUNDED BY THE EPA. TVA IS THE CONSTRUCTOR AND FACILITY OPERATOR. THE BECHTEL CORP. OF SAN FRANCISCO IS THE MAJOR CONTRACTOR, TEST DIRECTOR, AND REPORT WRITER.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 101 TENNESSEE VALLEY AUTHORITY SHAWNEE NO.10B 10 MW - RETROFIT COAL 2.9 PERCENT SULFUR CHEMICO LIME/LIMESTONE SCRUBBING STARTUP 4/72	REFER TO BACKGROUND INFORMATIION SECTION IN TABLE 3 OF THIS REPORT. THIS VENTURI/SPRAY TOWER LIME/LIMESTONE SCRUBBING SYSTEM HAS BEEN OPERATIONAL SINCE APRIL 1972. THIS TEST PROGRAM IS FUNDED BY THE EPA. TVA IS THE CONSTRUCTOR AND FACILITY OPERATOR. THE BECHTEL CORP. OF SAN FRANCISCO IS THE MAJOR CONTRACTOR,TEST DIRECTOR, AND REPORT WRITER.
I.D. NUMBER 102 TENNESSEE VALLEY AUTHORITY WIDOWS CREEK NO 8 550 MW - RETROFIT COAL 3.7 PERCENT SULFUR TENNESSEE VALLEY AUTHORITY LIMESTONE SCRUBBING STARTUP 2/77	CONTRACTS HAVE BEEN LET ON ALL MAJOR PROCESS EQUIPMENT INCLUDING THE SCRUBBER, MIST ELIMINATOR, PIPING, INSTRUMENTATION, HOT AIR INJECTION SYSTEM AND AIR HEATER FANS. THE ID FANS ARE NOW BEING INSTALLED AS IS THE LIMESTONE CRUSHER. THE LAST EQUIPMENT REQUISITION WILL BE COMPLETED BY JUNE 1976. THE SCHEDULED STARTUP DATE OF FEBRUARY 1977 IS STILL OPERATIVE.
I.D. NUMBER 103 TEXAS UTILITIES CO. MARTIN LAKE NO. 1 793 MW - NEW COAL 1.0 PERCENT SULFUR RESEARCH COTTRELL LIMESTONE SCRUBBING STARTUP 9/76	NEW SYSTEM, ERECTING BOILER. CONTRACT FOR FGD SYSTEM AWARDED. ENGINEERING DESIGN WORK IS PROCEEDING. FABRICATION OF THE TOWER HAS BEEN COMPLETED.
I.D. NUMBER 104 TEXAS UTILITIES CO. MARTIN LAKE NO. 2 793 MW - NEW COAL 1.0 PERCENT SULFUR RESEARCH COTTRELL LIMESTONE SCRUBBING STARTUP 6/77	NEW SYSTEM, ERECTING BOILER. CONTRACT FOR FGD SYSTEM AWARDED. ENGINEERING DESIGN WORK IS PROCEEDING. FABRICATION OF THE TOWER HAS BEEN COMPLETED.
I.D. NUMBER 105 TEXAS UTILITIES CO. MARTIN LAKE NO. 3 793 MW - NEW COAL 1.0 PERCENT SULFUR RESEARCH COTTRELL LIMESTONE SCRUBBING STARTUP 12/78	UTILITY HAS SIGNED A LETTER OF INTENT WITH RESEARCH COTTRELL. THE BOILER IS CURRENTLY BEING ERECTED AT THIS NEW FACILITY.

TABLE 2
STATUS OF FGD SYSTEMS DURING 12/75

COMPANY POWER STATION	CURRENT MONTH
I.D. NUMBER 106 TEXAS UTILITIES CO. MARTIN LAKE NO. 4 793 MW - NEW COAL 1.0 PERCENT SULFUR RESEARCH COTTRELL LIMESTONE SCRUBBING STARTUP 12/79	UTILITY HAS SIGNED A LETTER OF INTENT WITH RESEARCH COTTRELL. THE BOILER IS CURRENTLY BEING ERECTED AT THIS NEW FACILITY.
I.D. NUMBER 107 UNITED POWER ASSOCIATION COAL CREEK NO. 1 545 MW - NEW LIGNITE - 0.63 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 11/78	UTILITY IS INVESTIGATING BOTH REGENERABLE AND NON-REGENERABLE SYSTEMS.
I.D. NUMBER 108 UNITED POWER ASSOCIATION COAL CREEK NO. 2 545 MW - NEW LIGNITE - 0.63 PERCENT SULFUR NOT SELECTED NOT SELECTED STARTUP 11/79	UTILITY IS INVESTIGATING BOTH REGENERABLE AND NON-REGENERABLE SYSTEMS.
I.D. NUMBER 109 UTAH POWER & LIGHT CO. HUNTINGTON NO.1 415 MW - NEW COAL 0.5 PERCENT SULFUR CHEMICO LIME SCRUBBING STARTUP 6/77	A LETTER OF INTENT HAS BEEN SIGNED WITH CHEMICO. UNIT NO. 1 IS THE SECOND UNIT AT HUNTINGTON. UTAH REGULATIONS REQUIRE 80% SO2 REMOVAL.

STATUS OF FLUE GAS DESULFURIZATION AND SIMULTANEOUS REMOVAL OF SO_2 AND NO_x IN JAPAN

Jumpei Ando

Faculty of Science and Engineering
Chuo University
Kasuga, Bunkyo-ku, Tokyo

ABSTRACT

About 100 FGD plants including several large ones with a capacity of 250-500 MW went into operation in 1975. Operation of most plants has been trouble-free. The large plants have attained more than 97% availability. The total FGD capacity which was about 30 million Nm^3/hr (10,000MW equivalent) at the end of 1974 will exceed 70 million Nm^3/hr by the end of 1976. The rapid growth is due to the economical advantage of FGD over the use of low-sulfur fuels and to the reliability of plant operation. The growth, however, may slow down after 1977 for the following reasons:

(1) The ambient SO_2 concentration in big cities and industrial districts which was 0.05-0.09ppm in annual average several years ago has decreased to 0.02-0.03ppm, almost meeting the ambient standard.

(2) The recent economic depression has prevented industry from building new plants.

(3) A tendency of overproduction of FGD by-products has occurred alluring industry toward the use of low-sulfur fuels, because Japan has limited landspace available for discarding useless by-products.

(4) The stringent NO_x regulation has started to force industry to carry out flue gas denitrification. Several processes for simultaneous removal of NO_x and SO_2 have been developed and industry is waiting for the completion of new technology.

This paper will describe the recent status of FGD and simultaneous removal of SO_2 and NO_x.

Preceding page blank

STATUS OF FLUE GAS DESULFURIZATION AND SIMULTANEOUS REMOVAL OF SO_2 AND NO_x IN JAPAN

Jumpei Ando

1 Major waste gas desulfurization installations

Table 1 shows the major wet-lime/limestone process installations completed or to be completed between 1974 and 76. Most plants by-produce salable gypsum except Omuta plant, Mitsui Aluminum, using the Chemico-Mitsui process by-producing throw-away calcium sulfite sludge and two plants of Nippon Steel which use a convertor slag as the absorbent to by-produce low-grade gypsum for discarding.

Most of the new plants use limestone as the absorbent. Limestone is generally ground by wet mill to pass 325 mesh about 95% to recover more than 90% of SO_2. Operation parameters of major plants are listed in Table 2. The total capacity of wet-lime/limestone process plants will reach about 36 million Nm^3/hr (12,000MW equivalent) by the end of 1976 (Table 6).

Table 3 shows major plants with indirect lime/limestone processes including double alkali processes and similar processes using acidic absorbents such as dilute sulfuric acid and aluminum sulfate. Operation parameters of major plants are listed in Table 2. The more acidic the solution is, the less the SO_2 absorption capacity requiring the larger L/G ratio, the lesser the problem of scaling, and the easier the reaction with limestone. The total capacity of the indirect process plants will reach about 13 million Nm^3/hr (4,300MW) by the end of 1976.

Table 4 shows installations using other processes by-producing sulfuric acid, elemental sulfur, and ammonium sulfate. The total capacity is about 8 million Nm^3/hr. In addition, there are nearly 200 relatively small plants with sodium scrubbing, by-producing sodium sulfite for paper mills and sodium sulfate for waste or industrial use. The total capacity of these sodium scrubbing plants will reach about 14 million Nm^3/hr by the end of 1976 (Table 6).

Table 5 shows major processes classified by developers, the number of plants, and the total capacity. The combined capacity will reach 70 million Nm^3/hr (24,000MW) by the end of 1976. About one-half of the gas is from utility boilers and the rest from industrial boilers and other sources such as sulfuric acid and iron ore sintering plants.

Since most of the processes have been reported previously,$^{1,2)}$ the performance of major new plants for utility boilers and new developments in the steel industry will be given below.

Most plants purge wastewater as shown in Table 7. The amount of the water discharged from many of the plants is about the same as that from a "closed loop" in the U.S.A. which discards a calcium sulfite sludge containing about 50% water (or 0.5 water ratio). The wastewater is treated to reduce COD (chemical oxygen demand) usually to below 10mg/liter and suspended solids to about 30mg/liter.

Table 1 Major wet-lime/limestone process installations (1974-1976)

Process developer	User	Plant site	Absorbent	MW[a]	Type of plant[b]	Year of completion	Gypsum (tons/day)
Mitsubishi (MHI)	Kansai Electric	Kainan	CaO	150	Utility boiler	1974	20
"	Tokyo Electric	Yokosuka	$CaCO_3$	130	"	"	20
"	Kyushu Electric	Karita	CaO	175	"	"	70
"	Kawasaki Steel	Mizushima	"	232	Sintering plant	"	120
"	"	"	"	279	"	1975	90
"	"	Chiba	"	130	"	"	27
"	Kansai Electric	Amagasaki	"	125	Utility boiler	"	40
"	Mizushima Power	Mizushima	"	192	"	"	95
"	Kyushu Electric	Karatsu	$CaCO_3$	250 & 175	"	1976	110
"	"	Ainoura	"	250 x 2	"	"	190
"	Chugoku Electric	Owase	CaO	375 x 2	"	"	660
Mitsui-Chemico	Mitsui Aluminum	Omuta	$CaCO_3$	175[c]	Industrial boiler	1975	200
"	Elec. Pow. Dev.	Takasago	"	250[c]	Utility boiler	"	200
"	"	"	"	"	"	1976	200
Babcock-Hitachi	Chugoku Electric	Mizushima	"	104	"	1974	25
"	"	Tamashima	"	500 & 350	"	1975	350 & 240
"	Asahi Chemical	Mizushima	"	150	Industrial boiler	"	100
"	Kansai Electric	Osaka	"	160	Utility boiler	1976	40
Fuji Kasui-Sumitomo	Sumitomo Metal	Wakayama	"	123	Sintering plant	1975	40
"	"	Kashima	"	295	"	"	100
"	"	"	"	660	"	1976	220
"	"	Kokura	"	240	"	"	80
Chubu-MKK	Ishihara Sangyo	Yokkaichi	"	77	Industrial boiler	1974	50
Kobe Steel	Kobe Steel	Amagasaki	CaO	125	Sintering plant	1976	20
"	"	Nadahama	"	120	"	"	20
"	Nakayama Steel	Osaka	"	120	"	"	30
Nippon Steel	Nippon Steel	Wakamatsu	"	320	"	"	--
Chemico-IHI	Elec. Pow. Dev.	Isogo	$CaCO_3$	265[c]	Utility boiler	"	60
Kawasaki H.I.	Nippon Exlan	Saidaiji	"	83	Industrial boiler	1975	
"	Unitika	Okazaki	"	83	"	"	

a Actual for boilers and equivalent gas flow for others (at $3{,}000 Nm^3/hr$ per MW).

b Boilers are oil-fired unless otherwise noted.

c Coal-fired.

Table 2 Example of operation parameters of FGD plants by-producing gypsum and calcium sulfite

Process developer	Absorbent, precipitant (stoichiometry)	Capacity 1,000Nm^3/hr	Type of absorber	Slurry or solution		L/G liters/ Nm^3	Space velocity m/sec	Pressure drop* mm H_2O	SO_2 ppm		Moisture % of gypsum
				pH	conc.%				in	out	
Wet lime-limestone process											
MHI (Mitsubishi-JECCO)	CaO 0.95-1	550	GP[a)]	6.4	10	10	3.5	120	600	40	8-10
"	$CaCO_3$ 1	750	"	6	10	10	3.5	150	1,000	100	
Chemico-Mitsui	$Ca(OH)_2$ 1-1.05	385	Venturi	7	3-5	10-15		400	2,000	200	$CaSO_3$
Mitsui-Chemico	$CaCO_3$ 0.95-1	840	"	6	5	10-15		200	1,500	150	10-15
Babcock-Hitachi	$CaCO_3$ 1.1 - 1.2	1,460[c)]	pp[b)]	6.1	20	10	3.2	850	1,500	60	8-10
Fuji Kasui-Sumitomo	$CaCO_3$ 1.1 - 1.2	800	pp[b)]	6	5-6	5	4.5	200	500	20	
Chubu-MKK	$CaCO_3$	250	Screen	6	10	10	4	100	1,500	200	10-12
Ishikawajima-TCA	CaO	100	TCA		2	7	3		700	50	10-15
Kobe Steel	CaO[f)] 1.05	350	Spray	6-8	30	3	3	190	300	20	10
Indirect lime-limestone process											
Kureha-Kawasaki	Na_2SO_3, $CaCO_3$	1,260[d)]	GP[a)]	6.2	20	10	2.5	150	1,070	5	6-8
Showa Denko	Na_2SO_3, $CaCO_3$	500[c)]	Cone	6.8	25	2		250	1,400	40	8-10
Nippon Kokan	$(NH_4)_2SO_3$, CaO	150	Screen	6	30	2	3	250	700	30	8-10
Chiyoda	dil.H_2SO_4, $CaCO_3$	1,050	Tellerette	1	2-4	55-60	1		1,600	60	7-9
Kurabo	$(NH_4)_2SO_4$, CaO	100	Tellerette	4	10	6-10	2	100	1,500	80	8-10
Dowa	$Al_2(SO_4)_3$, $CaCO_3$	150	Tellerette	4	10	3[e)]	1.5	100	600	20	10-12
Kureha	CH_3COONa, $CaCO_3$	5	pp[b)]	5.5	20	7-8[g)]	2-2.5	280	1,400	1-3	6-7

a) Grid packed b) Perforated plate c) Four scrubbers in parallel d) Two scrubbers in parallel

e) For tail gas at 25°C. L/G 6-10 for flue gas. f) In $CaCl_2$ solution. g) Including limestone scrubbing.

* Including cooler, absorber and mist eliminator.

Table 3 Major indirect lime-limestone process installations completed in and after 1973 (double alkali type)

Process developer	Absorbent, precipitant	User	Plant site	MW[a]	Type of plant	Year of completion	Gypsum (tons/day)
Chiyoda	dil. H_2SO_4, $CaCO_3$	Daicel	Aboshi	31	Industrial boiler	1973	23
"	"	Mitsubishi Chem.	Yokkaichi	130	"	1974	60
"	"	Mitsubishi Pet. Chem.	Yokkaichi	230	"	1974	35
"	"	Hokuriku Electric	Toyama	250	Utility boiler	1974	180
"	"	Hokuriku Electric	Fukui	350	"	1975	290
"	"	"	Toyama	250	"	1976	180
Kureha-Kawasaki	Na_2SO_3, $CaCO_3$	Tohoku Electric	Shinsendai	150	"	1974	40
"	"	Shikoku Electric	Anan	450	"	1975	300
"	"	Shikoku Electric	Sakaide	450	"	1975	300
"	"	Kyushu Electric	Buzen	450x2	"	1977	500
Showa Denko	"	Showa Denko	Chiba	150	Industrial boiler	1973	110
"	"	Showa Pet. Chem.	Kawasaki	62	"	1974	70
"	"	Kanegafuchi Chem.	Takasago	93	"	1974	80
Showa Denko-Ebara	"	Nippon Mining	Saganoseki	37	H_2SO_4 plant	1973	$CaSO_3$
"	"	Poly Plastic	Fuji	65	Industrial boiler	1974	70
"	"	Kyowa Pet. Chem.	Yokkaichi	46	"	1974	35
"	"	Toho Zinc	Annaka	43	H_2SO_4 plant	1974	70
Tsukishima	Na_2SO_3, CaO	Kinuura Utility	Nagoya	63	Industrial boiler	1974	40
"	"	Daishuwa Paper	Fuji	85	"	1975	45
Kurabo Engineering	$(NH_4)_2SO_4$, CaO	Kurarey	Tamashima	31	"	1974	30
"	"	Daicel	Aboshi	53	"	1975	40
"	"	Jujo Paper	Ishinomaki	65	"	1975	40
Dowa Mining	$Al_2(SO_4)_3$, $CaCO_3$	Dowa Mining	Okayama	100	H_2SO_4 plant	1974	24
"	"	Naikai Engyo	Tamano	30	Industrial boiler	1976	30

a Actual for boilers and equivalent gas flow for other plants. Boilers are oil-fired.

Table 4 SO_2 removal installations which by-produce sulfuric acid and sulfur

Process developer	Absorbent	User	Plant site	MW[a]	Type of plant	Year of completion	By-product (tons/day)	
Wellman-MKK	Na_2SO_3	Japan Synth. Rubber	Chiba	70	Industrial boiler	1971	H_2SO_4	44
"	"	Chubu Electric	Nishinagoya	220	Utility boiler	1973	"	88
"	"	Japan Synth. Rubber	Yokkaichi	150	Industrial boiler	1974	"	44
"	"	Toyo Rayon	Nagoya	103	"	"	"	50
"	"	Mitsubishi Chem.	Mizushima	186	"	"	"	88
"	"	J. N. Railways	Kawasaki	217	"	1975	"	110
"	"	Kurashiki Rayon	Okayama	127	"	"	"	66
"	"	Shindaikyowa Oil	Yokkaichi	124	"	"	"	60
Wellman-SCEC	"	Toa Nenryo	Kawasaki	23	Claus furnace	1971	S	
"	"	Sumitomo Chem.	Chiba	120	Industrial boiler	1973	H_2SO_4	55
"	"	Fuji Film	Fuji	50	"	1974	Liquid SO_2	13
"	"	Sumitomo Chem.	Chiba	180	"	1975	H_2SO_4	88
"	"	Sumitomo Chem.	Niihama	50	"	"	"	25
Onahama-Tsukishima	MgO	Onahama Smelter	Onahama	28	Copper smelter	1972	"	240
Mitsui Mining	"	Mitsui Mining	Hibi	25	H_2SO_4 plant	1971	"	18
Chemico-Mitsui	"	Idemitsu Kosan	Chiba	162	Claus and boiler	1974	S	
Mitsui Mining	ZnO	Mitsui Mining	Kamioka	16	H_2SO_4 plant	1975	H_2SO_4	12
Sumitomo H.I.	Carbon	Kansai Electric	Sakai	53	Utility boiler	1971	"	17
Hitachi Ltd.	"	Tokyo Electric	Kashima	150	"	1972	H_2SO_4[b]	
"	"	Unitika	Uji	57	Industrial boiler	1975	H_2SO_4	
Nippon Kokan	$(NH_4)_2SO_3$	Nippon Kokan	Keihin	50	Sintering plant	1972	$(NH_4)_2SO_4$	
"	"	"	Fukuyama	253	"	1976	"	
Shell	CuO	Showa Yokkaichi	Yokkaichi	40	Industrial boiler	1973	S	
Mitsubishi-IFP	$(NH_4)_2SO_3$	Maruzen Oil	Shimozu	14	Claus furnace	1974	S	
TEC-IFP	"	Fuji Oil	Chiba	3	"	"	S	

[a] Actual for boilers and equivalent gas flow for others. Boilers are oil-fired.

[b] Reacted with limestone to produce gypsum.

Table 5 Major SO_2 recovery processes and plant capacities (Operational at end 1976)

Process	Absorbent	By-product	Number of plants	Total capacity $1,000Nm^3/hr$
Mitsubishi (MHI)	CaO, $CaCO_3$	Gypsum	31	17,908
Chiyoda	H_2SO_4-$CaCO_2$	Gypsum	15	5,488
Babcock-Hitachi	$CaCO_3$	Gypsum	9	5,341
Wellman-MKK	Na_2SO_3	H_2SO_4	10	4,120
Fuji Kasui-Sumitomo	CaO, $CaCO_3$	Gypsum	7	3,954
Oji	NaOH	Na_2SO_3	49	3,839
Kureha-Kawasaki	Na_2SO_3 - $CaCO_3$	Gypsum	3	2,940
IHI-TCA	NaOH	Na_2SO_3	26	2,787
Mitsui-Chemico	$CaCO_3$	Gypsum	5	2,739
Tsukishima-Bahco	NaOH	Na_2SO_3	20	2,655
Kurabo	NaOH	Na_2SO_3	70	2,580
Showa Denko-Ebara	Na_2SO_3-$CaCO_3$	Gypsum	16	2,458
Chemico-IHI	$CaCO_3$	"	2	1,800
Kureha	NaOH	Na_2SO_3	8	1,431
Wellman-SCEC	Na_2SO_3	H_2SO_4	6	1,298
Nippon Steel	Slag	Gypsum	2	1,170
Kobe Steel	CaO	"	3	1,070
Nippon Kokan	NH_3	$(NH_4)_2SO_4$	1	760
Tsukishima	Na_2SO_3-CaO	Gypsum	4	714
MKK	NaOH	Na_2SO_4	17	693
IHI-TCA	CaO, $CaCO_3$	Gypsum	3	650
Hitachi Ltd.	Carbon	H_2SO_4, Gypsum	2	590
Kurabo	$(NH_4)_2SO_4$-CaO	Gypsum	5	588
Kawasaki	CaO, $CaCO_3$, MgO	"	3	582
Chemico-Mitsui	MgO	$SO_2 \rightarrow S$	1	500
Dowa	$Al_2(SO_4)_3$-$CaCO_3$	Gypsum	4	430
Chemico-Mitsui	$Ca(OH)_2$	$CaSO_3$	1	385
Showa Denko	NaOH	Na_2SO_3	2	370
Chubu-MKK	$CaCO_3$	Gypsum	2	311
Nippon Kokan	CaO	"	2	227
"	$(NH_4)_2SO_3$-CaO	"	1	150
Sumitomo H.I.	Carbon	H_2SO_4	1	150
Shell	CuO	$SO_2 \longrightarrow S$	1	120
Onahama-Tsukishima	MgO	H_2SO_4	1	88
Mitsui Mining	"	"	1	80
Total			333	70.972

Table 6 Classification of major processes and plant capacities (operational at end 1976)

Classification of processes	Number of units	Capacity (X $10^6 Nm^3/hr$)*
Sodium sulfite recovery	192	14.3
Gypsum recovery		
Wet-lime/limestone	68	36.4
Indirect-lime/limestone (Double alkali type)	49	12.6
Sulfuric acid recovery	20	5.9
Other	4	1.8
Total	333	71.0

* $10^6 Nm^3/hr$ is equivalent to 330MW.

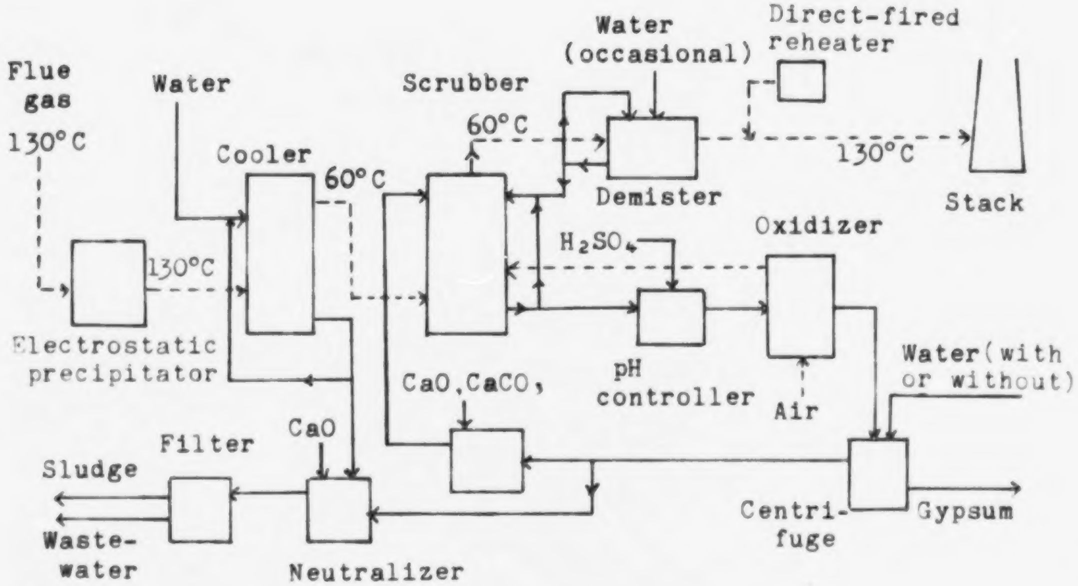

Flue gas
130°C
Water
Cooler
60°C
130°C
Electrostatic precipitator
Scrubber
60°C
Water (occasional)
Demister
Direct-fired reheater
130°C
Stack
H_2SO_4
Oxidizer
pH controller
Air
Water(with or without)
$CaO, CaCO_3$
CaO
Filter
Sludge
Waste-water
Neutralizer
Centri-fuge
Gypsum

Figure 1 Schematic flowsheet of wet lime/limestone gypsum process

Table 7 Wastewater from FGD plants

Process	User	Plant site	MW	Inlet SO_2 (ppm)	Waste-water (t/hr) (A)	Gypsum(t/hr)		Water ratio $\frac{(A+C)}{(A+B+C)}$	Waste-water (kg/MWhr)
						Solid(B)	Moisture(C)		
Mitsubishi (MHI)	Kansai Electric	Kainan	150	270	1.5	0.9	0.1	0.64	10
"	Kyushu Electric	Karita	188	600	3.7	2.2	0.2	0.64	20
"	Chubu Electric	Owase[a)]	750	1,480	14.0	29.0	2.9	0.37	19
Mitsui-Chemico	EPDC	Takasago[b)]	250	1,500	5.0	10.0	1.1	0.38	20
Babcock-Hitachi	Chugoku Electric	Mizushima	105	400	2.5	0.9	0.1	0.74	23
"	"	Tamashima	500	1,500	5.0	19.5	1.9	0.26	10
Chubu-MKK	Ishihara Chemical	Yokkaichi	85	1,300	3.5	2.2	0.2	0.62	41
Showa Denko	Showa Denko	Chiba	150	1,400	3.5	5.2	0.5	0.45	23
Chiyoda	Hokuriku Electric	Toyama	250	610	15.0	7.5	0.7	0.73	60
"	"	Fukui	350	1,540	24.0	14.0	1.4	0.64	68
Wellman-MKK	Chubu Electric	Nishinagoya	220	1,800	3.0	--	--	--	14

a) Designed value; the plant is under construction.

b) Coal-fired boiler. All others are for oil-fired boilers.

2 Performance of major new plants for utility boilers

2.1 Karita plant, Kyushu Electric

MHI lime-gypsum process (Mitsubishi-JECCO process)1)

The Karita plant with a capacity of treating 550,000Nm^3/hr flue gas from an oil-fired boiler (188MW equivalent) went into operation in November 1974. The plant is based on the one-absorber system (Figure 2) and uses lime. Operation parameters are listed in Table 2. Flue gas is first cooled to 55-60°C in a cooler and led into the absorber. The absorber inlet SO_2 concentration is about 500ppm and the outlet 20-30ppm. The load fluctuates between 550,000 and 300,000Nm^3/hr every day. The flow rate of the slurry is kept constant while the amount of lime is adjusted with the load.

The plant has been operated at 100% availability since its start-up except for the scheduled shutdown of the boiler from April 1 to May 14. On February 24, 1975, a flow rate adjusting bulb was stopped up but was repaired without interrupting the scrubber operation. In the inspection in April 1975 a considerable scaling was found on the mist eliminators. The eliminators had been washed with circulating liquor, but since May 1975 they have been washed alternately with fresh water and the liquor. Water balance is shown in Figure 2.

Two new units (175 and 250MW) by the MHI process using limestone at Karatsu station, Kyushu Electric, has started operation recently. Designed operation parameters are shown in Table 2.

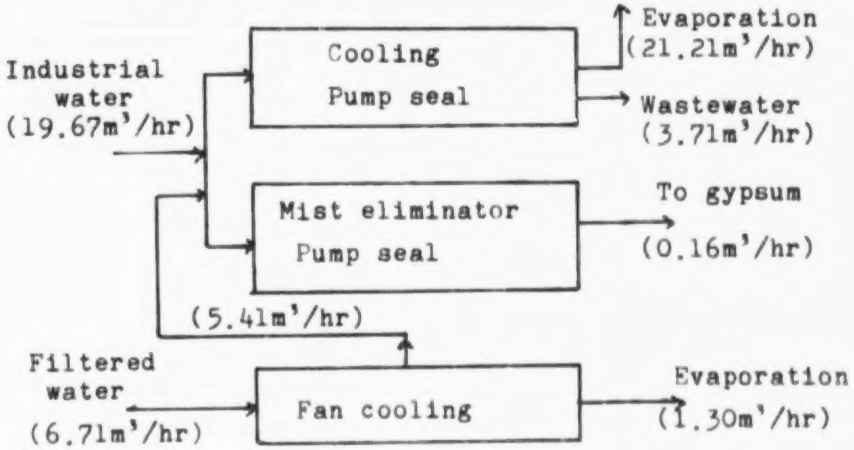

Industrial water
(19.67m³/hr)
Cooling
Pump seal
Evaporation
(21.21m³/hr)
Wastewater
(3.71m³/hr)
Mist eliminator
Pump seal
To gypsum
(0.16m³/hr)
(5.41m³/hr)
Filtered water
(6.71m³/hr)
Fan cooling
Evaporation
(1.30m³/hr)

Figure 2 Water balance (Karita plant, MHI process)

2.2 Takasago plant, Electric Power Development Co.

Mitsui-Chemico limestone-gypsum process (Mitsui Miike process)[1,2]

This plant is based on the Mitsui-Chemico process developed by Mitsui Miike Machinery Co. It is the first Mitsui-Chemico process unit to be used commercially for a coal-fired utility boiler, and has a capacity of treating 840,000Nm^3/hr flue gas from a 250MW boiler. The plant consists of two single-stage Chemico scrubbers placed in series, a pH controller to reduce the pH of the calcium sulfite slurry by introducing the flue gas, and two reactors for oxidation of the sulfite to gypsum by air. A catalyst is used to promote both SO_2 removal and oxidation. The flue gas passed through an electrostatic precipitator containing 1,500ppm SO_2 and 80mg/Nm^3 dust is treated to remove 90% of the SO_2 and 75% of the dust. Limestone, 95% under 325 mesh, is used.

The plant went into operation in January 1975 and has been in smooth operation except for several short shutdowns to clean up the mist eliminators. The eliminators have been washed with the circulating liquor and fresh water. The availability reaches 98% except in periods of boiler shutdown. The load fluctuates between 170 and 250MW every day. The slurry flow rate in the scrubbers is kept constant.

The daily requirements for the operation are as follows:

$CaCO_3$ 118 tons   Power 136,000kWhr   Oil 41 liters(reheating)
Steam 13 tons (heating of oil)   Industrial water 1,200 tons
Catalyst $350

The coal contains about 500ppm chlorine. In order to keep the chlorine concentration of the scrubber liquor below 7,000ppm, about 5 tons/hr of wastewater is purged. The relationship between the amount of purge water and the chlorine concentration of coal and equilibrated scrubber liquor is shown in Figure 3.

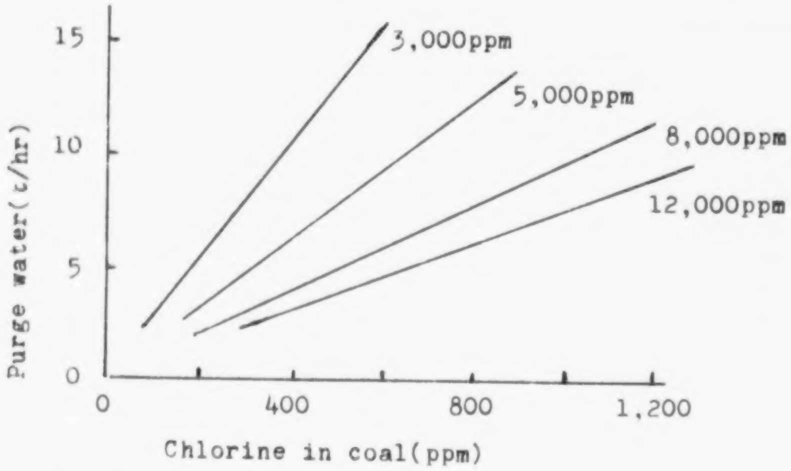


Figure 3 Purge water and chlorine concentration

Another plant with the same capacity is near completion at Takasago. The new plant has one reactor in place of the two reactors of the present plant.

2.3 Tamashima plant, Chugoku Electric

Babcock-Hitachi process[1, 2)]

Hitachi Ltd. constructed a large plant (500MW full scale) using Babcock-Wilcox scrubbers and an oxidizing system to by-produce gypsum. Four scrubbers were installed of which three are is use and one is for stand-by. The plant has been in operation since July 1975. Flue gas from an oil-fired boiler containing 1,500ppm SO_2 is first cooled to 55°C in venturi scrubbers and then led into absorbers with 6 stages of perforated plates where about 95% of SO_2 is removed. The pressure drop is fairly heavy--230mm H_2O in the venturi and 600mm in the absorber. A new type of mist eliminators designed by Hitachi are used.

The plant has been used for base-load operation. Power consumption reaches 3.6% of the power generated. Operation has been trouble-free except for minor erosion and scaling problems. Availability of the plant has reached about 97%.

About 93 tons/hr of industrial water is used of which 13 tons are for washing the mist eliminator. At the beginning of the operation, about 5 tons/hr wastewater was purged to keep the chlorine concentration of the circulating liquor below 1,000ppm. The amount of wastewater has been substantially reduced recently. Another plant (350MW) at Tamashima has been in operation since early 1976.

2.4 Sakaide plant, Shikoku Electric

Kureha-Kawasaki sodium limestone process

The Sakaide plant with a capacity of treating 1,260,000Nm^3/hr flue gas containing 1,050ppm SO_2 from a 450MW oil-fired boiler went in operation in August 1975. The process is similar to that for Shinsendai plant, Tohoku Electric[1, 2)] except that the Sakaide plant has a unit to remove magnesium from the circulating liquor (magnesium tends to delay the reaction of limestone with sodium sulfite) and that gypsum is washed with water to reduce sodium, to make it available for wall-board production. Operation parameters are shown in Table 2. More than 99% of SO_2 is removed by sodium sulfite scrubbing. Operation has been smooth since start-up. Load fluctuates between 1,260,000 and 534,000Nm^3/hr every day. The sodium sulfate concentration of the circulating liquor has reached 11-12% exceeding the designed value of 8% but a high SO_2 recovery ratio has been attained.

This plant is characterized by the absence of water purging from the system. At full-load operation, 63.5 tons/hr water is charged of which 29.9 tons are used for gypsum wash. The same amount, 63.5 tons, leaves the system, of which 62.0 tons are evaporated and the rest is contained in the by-product gypsum. Chlorine concentration of the circulating liquor has reached 3,500ppm but has caused no corrosion problem because the liquor contains little oxygen which tends to cause stress corrosion in the presence of much chlorine. In the Kureha-Kawasaki process, oxidation is carried out with the calcium sulfite separated from the mother liquor and therefore the oxygen content of the liquor is kept low. It seems that the chlorine input from fuel and water and output with gypsum containing 7-9% moisture have been nearly equalized at the 3,500ppm concentration.

2.5 Fukui plant, Hokuriku Electric

Chiyoda dilute sulfuric acid limestone process[1)]

The Fukui plant has been in smooth operation since its start-up in summer 1975 treating 1,050,000Nm^3/hr flue gas from a 350MW oil-fired boiler. The plant has a double cylinder type absorber-oxidizer. Operation parameters are listed in Table 2. The requirements are 150 tons/day limestone, 3.8m^3/hr oil for reheating, 11.8MW electric power, and 2,000 tons/day industrial water. About 580 tons/day wastewater is purged after being treated to keep the pH at 7.9, suspended solids below 5 mg/liter and COD below 10 mg/liter. Plant operation is easy and requires only 2 operators per shift.

3 New developments in SO_2 removal processes for steel industry

A major SO_2 source other than boilers is iron-ore sintering plants, which emit 300,000-1,000,000Nm^3/hr waste gas per plant containing 400-600ppm SO_2 and much dust. Several plants for desulfurization of the gas has been constructed by MHI since 1972 and are in successful operation. Recently three processes have been developed mainly for the treatment of the gas. New developments will be described below.

3.1 Fuji Kasui-Sumitomo limestone-gypsum process[1)]

Fuji Kasui Kogyo has developed a limestone-gypsum process jointly with Sumitomo Metal using the Moretana scrubber and recently completed two plants for Sumitomo. One is Wakayama plant (375,000Nm^3/hr) and the other Kashima plant (800,000Nm^3/hr). Operation parameters are shown in Table 2.

The Moretana scrubber is fitted with four perforated plates made of stainless steel. The holes, L/G ratio, and the gas velocity are specially designed to give extreme turbulence, producing foam layers 400 to 500mm thick which result in a high SO_2 recovery ratio. An oxidizer developed by Fuji Kasui is used. Other facilities and systems are similar to those of other Japanese processes (Figure 1).

The Wakayama plant started operation in May 1975 and has been in smooth operation except for a defect in the plastic lining in a cooler which was found at the beginning of the operation and was repaired. Availability of the plant is 98% except for the scheduled shutdown of the sintering plant that normally occurs one day in about two months. The mist eliminator is washed intermittently (once in 30 minutes) with the circulating liquor and fresh water alternatively. The pressure drop in the mist eliminator which is 30mm H_2O at the beginning gradually increases while it is washed with the circulating liquor. When the pressure drop reaches 50mm, fresh water is used in place of the liquor until the pressure drops back to 30mm. The ratio of liquor to fresh water is about 80 to 20.

The Kashima plant started operation in September 1975 and has been operated at 100% availability.

3.2 Nippon Steel slag process (SSD process)

Nippon Steel has developed a process which uses convertor slag as the absorbent. The slag contains about 40% CaO, 16% SiO_2, 3% MgO, 3% Al_2O_3, and 35% FeO and Fe_2O_3 and is useless. Nippon Steel has operated a prototype plant with a capacity of treating 200,000Nm^3/hr waste gas from a sintering plant since 1974 and is now constructing a commercial plant (1,000,000Nm^3/hr) which will come on-stream in 1976.

The process is similar to other Japanese lime/limestone-gypsum processes except that it uses no oxidizer. The gas is cooled and led into two absorbers in series to remove 95% of SO_2. The slag is fed to the second absorber to produce a calcium sulfite slurry which is led to the first scrubber and entirely oxidized into gypsum in the scrubber due to a low pH and the presence of much iron compounds which act as a catalyst. The by-product gypsum contains about 40%

impurities and has been discarded. There has been some scaling problem to be solved to ensure a long-term continuous operation. The process may be useful for steel producers who normally have large amounts of useless slag.

3.3 Kobe Steel calcium chloride process (Cal process)$^{1)}$

Kobe Steel has developed a new process using a 30% calcium chloride solution dissolving lime as the absorbent. A pilot plant (50,000Nm^3/hr) has been operated and two commercial plants (Table 1) have just come on-stream to treat waste gas from iron ore sintering plants.

Calcium chloride solution dissolves 6-7 times as much lime as does water. High SO_2 recovery is attained at a low L/G (Table 2). The flowsheet is similar to that in Figure 1 except that calcium sulfite discharged from the scrubber is separated from most of the mother liquor, reslurried with water, and sent to an oxidizer and that the liquor from the gypsum centrifuge is sent to a cooler giving no wastewater at all. Since vapor pressure of the liquor is low, the temperature of the gas after the scrubbing reaches 70°C as compared with the 55-60°C for the usual wet process and thus less energy for reheating is required.

The mist eliminator is washed with the circulating liquor. The solubility of gypsum in the liquor is very low (nearly 1/100 of that in water) and the evaporation of the liquor does not cause scaling.

Continuous operation of the pilot plant for about 6 months showed that a soft deposit formed on the wall of the absorber when the L/G ratio was smaller than 1 but the deposit could be washed off by using an L/G larger than 2. A highly corrosion-resistant material is required for the cooler. Operation of commercial plants will enable further evaluation of the process.

4 Simultaneous removal of SO_2 and NO_x

4.1 Outline

NO_x removal technology has been developed in Japan remarkably since 1973 when a stringent ambient standard for NO_2 (0.02ppm in daily average) was set forth. Among many processes, selective catalytic reduction of NO_x by ammonia at about 400°C has been considered most feasible. Several commercial plants treating flue gas from oil-fired industrial boilers (200,000-450,000Nm^3/hr) are in operation and many plants are under construction with the reduction process. This dry denitrification process, however, is not suited for use in conjunction with FGD (for which a wet process is economical), because it will require a large heat exchanger and a considerable amount of energy for heating (No. 3, Figure 4) or an expensive hot electrostatic precipitator (No. 4, Figure 4). Under this situation, many dry and wet processes for simultaneous removal of SO_2 and NO_x (No. 1 and 2, Figure 4) have been developed and many plants are in operation as shown in Table 8.

4.2 Oxidation reduction process

One of the most promising ways for simultaneous removal is the reduction of NO_2 into N_2 using sulfite ion formed by SO_2 absorption.

$$4SO_3^{--} + 2NO_2 = 4SO_4^{--} + N_2$$

Since NO is less reactive, it must be oxidized into NO_2 first. As the oxidizing agent, ClO_2 is used by Fuji Kasui and Ozone is used by Chiyoda and MHI. Ozone is a good oxidizing agent but is costly. To convert NO in flue gas from a 350MW boiler to NO_2, 400-600kg/hr ($500-700/hr) of ozone is required assuming that the gas contains 200-300ppm NO. NO concentration in flue gas should be kept low by combustion control. Chlorine dioxide (ClO_2) is also very effective and converts NO to NO_2 within a second. Although it adds a chloride to the by-product and complicates treatment, the cost is about one-half that of ozone.

Fuji Kasui has constructed three plants using the Moretana scrubber (Table 8). Flue gas containing 1,000-1,500ppm SO_2 and 200-300ppm NO_x is cooled to 55-60°C, mixed with equimolecular ClO_2, and subjected to sodium scrubbing to remove about 90% of NO_x and 98% of SO_2. About one-half of the removed NO_x is converted to N_2 and the rest forms sodium nitrate and nitrite.3) The liquor from the scrubber is evaporated to recover a high-purity sodium sulfate for chemical use. The rest of the by-product has to be discarded.

Chiyoda has developed a simultaneous removal process by simply adding an ozone-oxidation step before the standard FGD process. Together with more than 90% of SO_2, more than 60% of NO_x is removed, forming N_2O and some calcium nitrate.3) Purge liquor is sent to a wastewater treatment system.

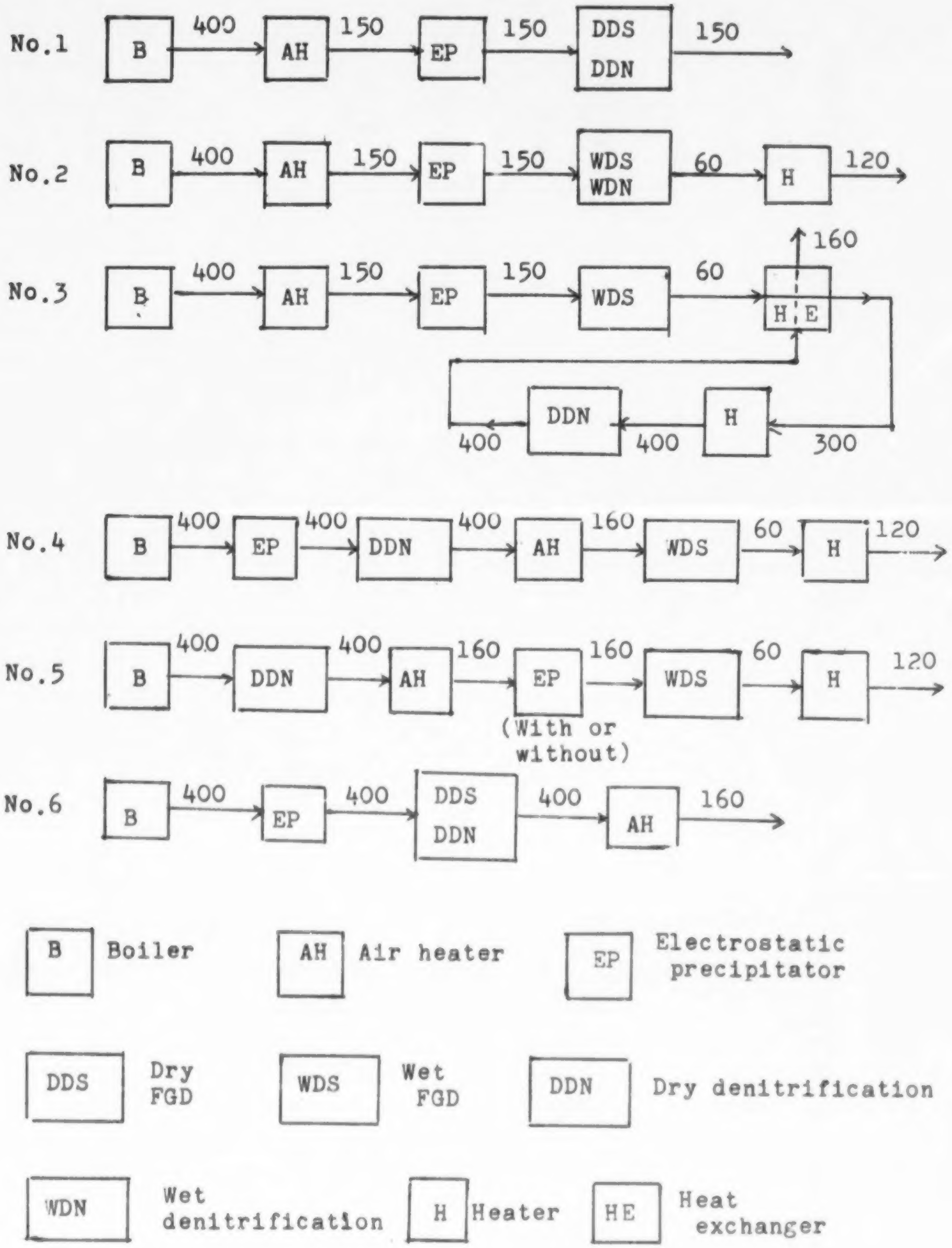

No.1
B
400
AH
150
EP
150
DDS
DDN
150
No.2
B
400
AH
150
EP
150
WDS
WDN
60
H
120
No.3
B
400
AH
150
EP
150
WDS
60
H
E
160
DDN
H
400
400
300
No.4
B
400
EP
400
DDN
400
AH
160
WDS
60
H
120
No.5
B
400
DDN
400
AH
160
EP
(With or without)
160
WDS
60
H
120
No.6
B
400
EP
400
DDS
DDN
400
AH
160
B
Boiler
AH
Air heater
EP
Electrostatic precipitator
DDS
Dry FGD
WDS
Wet FGD
DDN
Dry denitrification
WDN
Wet denitrification
H
Heater
HE
Heat exchanger

Figure 4 Models of combination of FGD and denitrification
(Figures show gas temperature)

Table 8 Installations for simultaneous removal of SO_2 and NO_x

Process developer	Type of process	Plant owner	Plant site	Capacity Nm^3/hr	Source of gas	Completion	By-product
Wet process							
(Fuji Kasui - Sumitomo Metal)	(Oxidation reduction)	Sumitomo Metal	Amagasaki	62,000	Boiler[a)]	Dec. 73	($NaNO_3$, NaCl Na_2SO_4)
"	"	Toshin Steel	Fuji	100,000	Furnace[b)]	Dec. 74	"
"	"	Sumitomo Metal	Osaka	39,000	Boiler[a)]	Dec. 74	"
Chiyoda	"	Chiyoda	Kawasaki	1,000	"	1973	(Gypsum $Ca(NO_3)_2$)
Mitsubishi H.I.	"	Mitsubishi H.I.	Hiroshima	2,000	"	Dec. 74	Gypsum, N_2
Ishikawajima H. I.	"	Ishikawajima	Yokohama	5,000	"	Sep. 75	"
Kobe Steel	"	Kobe Steel	Kakogawa	1,000	Furnace[d)]	Dec. 73	Gypsum, N_2
"	"	"	"	50,000	"	Mar. 76	"
Kawasaki H.I.	(Magnesium scrubbing)	EPDC	Takehara	5,000	Boiler[c)]	Dec. 75	(Gypsum $Ca(NO_3)_2$)
Kureha Chemical	Reduction	Kureha Chem.	Nishiki	5,000	Boiler[a)]	Apr. 75	Gypsum, N_2
Chisso Engineering	"	Chisso P.C.	Goi	300	Boiler[a)]	1974	$(NH_4)_2SO_4$
Mitsui S.B.	"	Mitsui P.C.	Chiba	150	"	"	H_2SO_4, N_2
Dry process							
Shell	CuO, NH_3	Showa Y.S.	Yokkaichi	120,000	"	1975	$SO_2 \rightarrow S$, N_2
Unitika	Carbon, NH_3	Union Glass	Hirakata	4,500	Furnace[e)]	"	H_2SO_4, N_2
Sumitomo H.I.	"	Sumitomo Power	Niihama	150	Boiler[a)]	1974	"
Ebara	Electron beam	Ebara	Fujisawa	1,000	"	"	Mist, dust

a) Oil-fired boiler b) Metal heating furnace c) Coal-fired boiler

d) Iron ore sintering furnace e) Glass melting furnace

MHI has developed a simultaneous removal process based on the lime/limestone gypsum process. Ozone is used for oxidation and a catalyst is used to convert virtually all of the removed NO_x to N_2. More than 70% of NO_x is removed with more than 90% of SO_2. A portion of the circulating liquor is treated to remove certain compounds formed by a side reaction.

Ishikawajima-Harima Heavy Industries (IHI) has also developed a simultaneous removal process based on wet-lime/limestone scrubbing using ozone.5)

Kobe Steel has developed a process using a calcium chloride solution containing $Ca(OCl)_2$ as an oxidizing and neutralizing agent.3) Following tests with a pilot plant (1,000Nm^3/hr), a larger pilot plant (50,000Nm^3/hr) is under construction. The by-products are gypsum and N_2 or HNO_3.

4.3 Other wet simultaneous removal processes

Kawasaki Heavy Industries has operated a pilot plant with a capacity of treating 5,000Nm^3/hr flue gas from a coal-fired boiler using magnesium scrubbing and lime addition.3) Magnesium sulfate and calcium nitrate are formed and are reacted to precipitate gypsum. Calcium nitrate is obtained as a by-product.

Chisso Eng. has operated a pilot plant in which SO_2 and NO_x are absorbed with an ammoniacal solution containing a catalyst to reduce the absorbed NO_x into NH_3 by ammonium sulfite and bisulfite to by-produce ammonium sulfate.3) The overall reaction may simply be expressed in the following way:

$$2NO + 5SO_2 + 8NH_3 + 8H_2O = 5(NH_4)_2SO_4$$

Kureha Chemical has modified the sodium acetate-limestone FGD process for simultaneous removal to by-produce gypsum and N_2.3) Although the total system is fairly complex, an intermediate compound--sodium imidodisulfonate, $NH(SO_3Na)_2$--has been found useful as a substitute for sodium tripolyphosphate which has been used in large amounts as a builder for detergents and has caused environmental problems. Commercial use of the imidodisulfonate would make the Kureha process promising.

Mitsui Shipbuilding has recently developed a process that uses a solution containing ferrous ion and EDTA to absorb SO_2 and NO_x. The liquor after the absorption is sent to a reducing step to convert the ferric ion formed by oxygen in flue gas back to ferrous ion. Most of the reduced liquor is returned to the absorption step; the rest is sent to a stripper which generates concentrated SO_2 and NO, which are sent to a sulfuric acid plant. The NO may be reduced to N_2 by a conventional process or used for nitric acid production.

4.4 Activated carbon process

Takeda Chemical has produced an activated carbon containing metallic components to promote the reaction of NO_x with ammonia to form N_2. Higher temperature is favorable to the reaction but decreases the SO_2 absorbing capacity (Figure 5). Optimum temperature for simultaneous removal by this process is about 250°C.

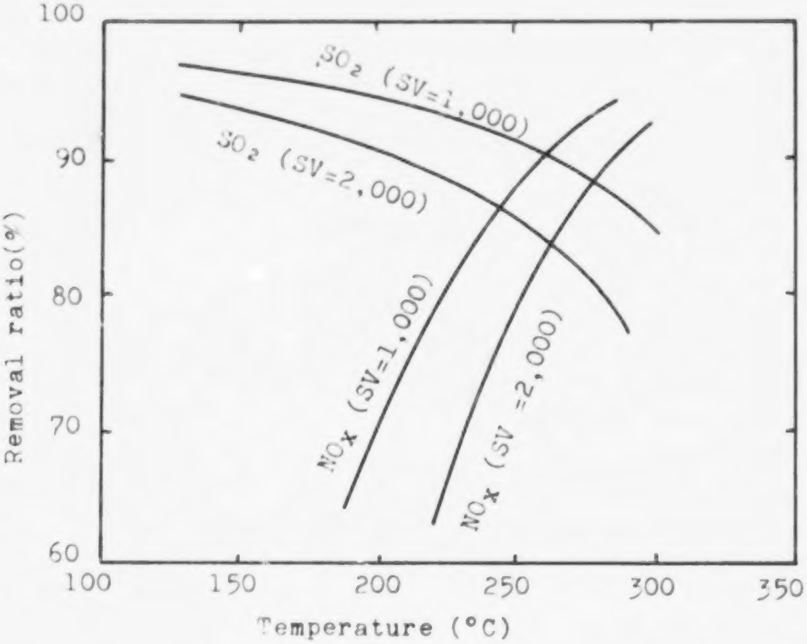


Figure 5 Schematic diagram of SO_2 and NO_x removal by activated carbon at different temperatures and SVs.

Unitika Co. recently started operation of a pilot plant with a capacity of treating 4,500Nm^3/hr flue gas from a glass melting furnace. About 90% of NO_x and SO_2 is removed. The carbon that has absorbed SO_2 is heated to 350°C in a reducing hot gas to release concentrated SO_2 for sulfuric acid production. Ammonium sulfate and sulfite which tend to form on the carbon are decomposed to SO_2 and N_2 in the regeneration step.

Sumitomo Heavy Industries (former Sumitomo Shipbuilding) has developed a similar process using moving beds of carbon.

Hitachi Ltd. has found that activated carbon treated with ammonium bromide is effective even at 100°C for NO_x reduction by ammonia.4) The low temperature activity may result in energy saving (No. 1, Figure 4).

Dust, ammonium sulfate and sulfite which tend to deposit on the carbon, however, present a problem for carbon processes.

4.5 Other dry simultaneous removal processes

Copper oxide used as an absorbent of SO_2 in the Shell process works as catalyst in the reaction of NO_x with ammonia. Yokkaichi plant, SYS, treating 120,000Nm^3/hr flue gas from an oil-fired boiler by the Shell process has introduced ammonia into a reactor at 400°C since 1975. Up to about 70% of NO_x can be removed. Copper sulfate formed by SO_2 absorption is reacted with hydrogen to generate concentrated SO_2, which is sent to a Claus furnace to produce sulfur.

Ebara Manufacturing Co. has developed a unique process for simultaneous removal by electron beam radiation.3) Flue gas is introduced in a reactor and exposed to the beam. About 80% of SO_2 and 90% of NO_x can be removed, forming a sulfuric acid mist and a powdery product containing S, N, O and H, which are caught by an electrostatic precipitator. Investment cost and power consumption seem high.

5. Economic aspects and by-products

5.1 Flue gas desulfurization

Investment cost in battery limits for plants (150-300MW) using wet and indirect lime/limestone processes by-producing gypsum was $20-30/kW in 1973, increased to $70-100/kW after the oil shock, and recently decreased to $50-80/kW. Plants by-producing sulfuric acid cost 20-70% more. The present desulfurization cost by lime/limestone processes including a depreciation for seven years ranges from $18 to $22/kiloliter oil or 4 to 5 mil/kWhr at 6,000 hours annual operation. The current price of a kiloliter of heavy oil is about $75 for high-sulfur oil (S=3%) and $100 for low-sulfur oil (S=0.3%). The economic advantage of FGD over the use of low-sulfur fuels and the reliability of FGD processes encouraged industry and power companies to construct many plants. But the plant building rush subsided recently with the oversupply of by-products.

In addition to the rapid increase of by-products as shown in Figure 6, the recent economic depression in Japan which resulted in the decrease in the demand has enlarged the oversupply (Figure 8). Last year, the total inventory of by-product gypsum at phosphoric acid and FGD plants exceeded 1 million tons--which is very little compared with that in Florida, U.S.A., but presented a problem in Japan which has a limitation in landspace. Efforts have been made to develop new uses for gypsum particularly in the field of new construction materials.

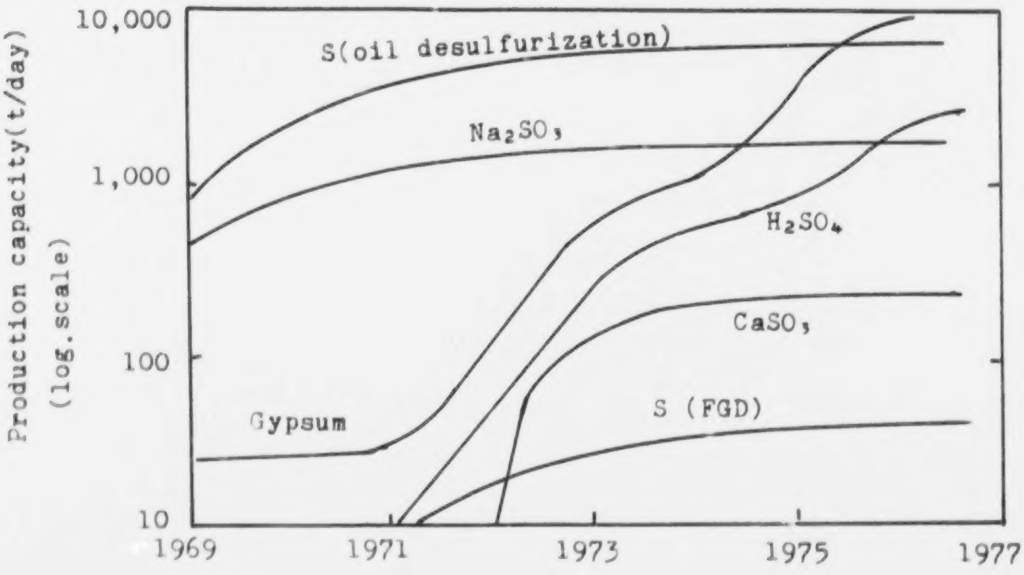


Figure 6 Production capacity of desulfurization

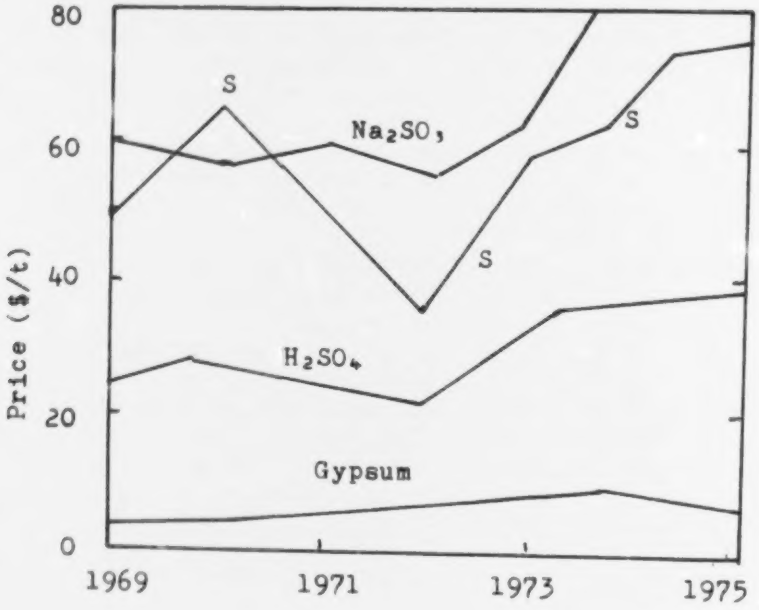
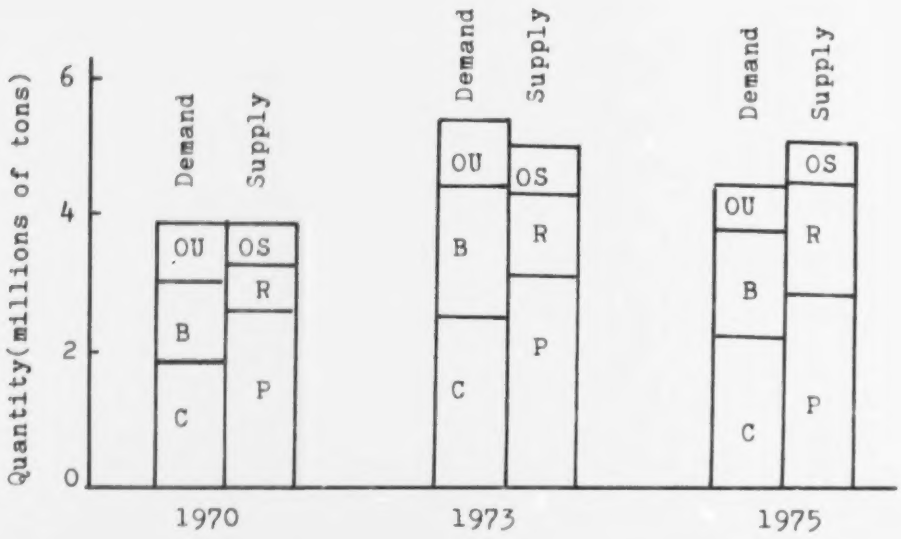


Figure 7 Price of by-products



Demand: C:Cement B:Board OU:Other uses
Supply: P:Phosphogypsum R:Recovered OS:Other sources

Figure 8 Demand for and supply of gypsum in Japan

The by-product gypsum from flue gas from coal-fired boilers at Omuta plant, Mitsui Aluminum and Takasago plant, E.P.D.C. contains less than about 6% impurities (fly ash, limestone, etc.) and has been used mainly for cement and partially for wallboard. Isogo plant, E.P.D.C. and Wakamatsu plant, Nippon Steel, both under construction, will produce gypsum with more than 20% impurities to be discarded. The first unit of Mitsui Aluminum in operation since 1972 by-producing calcium sulfite sludge will have to be modified to by-produce gypsum because the sludge pond is nearly full. Although gypsum is in oversupply, throw-away calcium sulfate sludge will not increase because the sludge needs much more land space than does gypsum.

Elemental sulfur is a desirable by-product but FGD processes by-producing sulfur seem very expensive except for oil refineries that have Claus furnaces. To by-produce elemental sulfur, hydrodesulfurization of heavy oil may be more economical while it is more expensive than FGD by the wet lime/limestone process (Table 9).

Table 9 Cost comparison at various desulfurization ratios

($/kl, at 6,000 hours annual operation)

	70%	80%	90%	97%
FGD (wet-lime/limestone process)	17	18	19	20
Hydrodesulfurization of oil	17	19	22	27

Two commercial hydrodesulfurization plants recently started operation to reduce sulfur to below 0.3% (over 90% removal) using heavy oil with little metallic impurities which tend to poison the catalyst. Most heavy oils are rich in impurities and difficult to desulfurize over 80%. For such heavy oils, vacuum distillation followed by hydrodesulfurization of the distillate and decomposition of the residue with partial gasification has been considered rational. The first commercial plant with the decomposition started operation in January 1976 using the Kureha process (or Eureka process). The second plant is near completion using the Flexicoking process.

5.2 Simultaneous removal

Fuji Kasui claims that the plant cost for simultaneous removal ranges from $60 to 90/kW and the operation cost is roughly $30/kl oil including depreciation for 7 years. The costs for other processes are uncertain because those are still in a test stage.

Activated carbon processes have an advantage in that they are a dry process and can be operated at relatively low temperatures. Carbon for simultaneous removal is fairly expensive at present, viz., about $8,000/t, while carbon for FGD costs about $3,000/t. To treat flue gas from a 350MW boiler at a SV of 1,000, about 1,000 tons of carbon is required. Production of much cheaper activated carbon is desired.

The Shell process seems costly for FGD but its capability of simultaneous removal may compensate for the disadvantage.

Among the by-products from wet processes, sodium nitrate and nitrite have little use. Calcium nitrate may be used as fertilizer but its demand is limited due to the low nitrogen content and high hygroscopicity. Ammonium nitrate and nitric acid are better by-products.

By the hydrodesulfurization of heavy oil, 30-40% of nitrogen in the oil is also removed resulting in 10-20% abatement of NO_x in flue gas. Tests have been made to remove much more nitrogen in the oil along with sulfur. To compete with possible low-sulfur low-nitrogen oil, the operation cost of simultaneous removal from flue gas should be less than about $30/kl and the process must not have difficulty in wastewater treatment and disposal of by-products.

Efforts for denitrification in Japan has been exerted mainly for flue gas from oil burning containing 200-300ppm NO_x because oil is the major fuel in the country. The need for denitrification may be larger for flue gas from coal burning which normally contains more than 600ppm NO_x which is not easy to reduce to below 400ppm by combustion modifications. Denitrification will become more and more important in many countries as the consumption of fossil fuels increases.

Conversion figures

1 m^3 = 35.3 cubic feet

1 liter = 0.26 gallon

1 kl = 6.29 barrels

1 Nm^3/hr = 0.59 scfm

L/G 1 liter/Nm^3 = 7.4 gallon/1,000scf

References

1) J. Ando, Recent Developments in Desulfurization in Japan (January 1975). Being published by U.S. EPA through PEDCO.

2) G. A. Hollinden and F. T. Princiotta, Sulfur Oxide Control Technology, Visits in Japan (March 1974). Interagency Technical Committee.

3) J. Ando and H. Tohata, NO_x Abatement Technology in Japan for Stationary Sources (March 1975). Being published by U.S. EPA through PEDCO.

4) M. Seki, Y. Sakurai and K. Yoshida, Ammonium Halide-Activated Carbon Catalyst to Decompose NO_x in Stack Gases at Low Temperatures around 100°C. American Chemical Society (August 1975, at Chicago).

5) S. Yamada, Y. Watanabe and H. Uchiyama, Bench-Scale Tests on Simultaneous Removal of SO_x and NO_x by wet Lime and Gypsum Process Ishikawajima-Harima Engineering Review, Jan. 1976, Vol.16, No.1. (In Japanese)

FLUE GAS DESULFURIZATION ECONOMICS

G. G. McGlamery, H. L. Faucett, R. L. Torstrick, and L. J. Henson

Office of Agricultural and Chemical Development
Tennessee Valley Authority
Muscle Shoals, Alabama

ABSTRACT

Described herein are several phases of flue gas desulfurization (FGD) process and byproduct economic evaluations being carried out by the TVA Office of Agricultural and Chemical Development for EPA and the TVA Office of Power. Included are recently updated investment (mid-1977) and annual revenue requirements (mid-1978) for five leading FGD processes evaluated by TVA in 1974. Also included are results from a cost sensitivity study of limestone scrubbing options based on data from the EPA-Bechtel Corporation-TVA test program at Shawnee, plus some preliminary evaluations of promising double-alkali and regenerable processes.

In addition to these results, several new or expanded evaluation programs are described. Included are:

1. A computer program prepared by Bechtel and TVA to screen process options of lime-limestone scrubbing using Shawnee data.
2. Marketing studies covering volume, pricing, and location aspects of various FGD byproducts.
3. Evaluations of advanced FGD processes, NO_x control processes, and systems for clean fuel from coal.
4. A sludge disposal system design and cost study.
5. Process energy optimization studies.

The status of these programs is also discussed; some phases of these programs are nearing completion, while others are either just getting under way or yet to be started.

FLUE GAS DESULFURIZATION ECONOMICS

INTRODUCTION

Throughout the developmental stages of flue gas desulfurization (FGD) technology, process economics has been one of the more interesting and debatable subjects, to say the least. From the earliest cost estimates of the late 1960's to now, and no doubt in the future, the estimated and actual costs of such systems have been increasing to higher and higher levels. But, so has the cost of most everything else in today's inflation-ridden society. Nevertheless, not all of the increases in cost estimates for FGD systems can be attributed to inflation, since like most new emerging technologies, as experience and knowledge increase in the early years of application, so does the expected real costs of such facilities.

Some perspective can be gained from Figures 1 and 2 which show the capital cost of the generalized limestone slurry scrubbing process as estimated by TVA over the past decade. Although the scope of these estimates has tended to change slightly during this period, the increases are due to both improvements in knowledge of process requirements and inflated labor and equipment costs.

By assuming a relatively simple example case, we have been fairly successful in assessing the comparative economics of FGD systems. However, on a total magnitude basis, it has been much more difficult to predict representative costs of specific systems. For that matter, estimators throughout the industry have encountered this difficulty, primarily for the following reasons.

1. Variable project scope, time frame, design, and operating conditions.
2. Numerous process options.
3. Vendor optimism and user skepticism.
4. Confusion over regulatory requirements.
5. Demanding physical and chemical environment.
6. Variable byproducts and byproduct value/disposal cost.
7. Shortage of actual design and operating data.

At least for the near future, we expect most of these reasons to remain as impediments to total cost assessment.

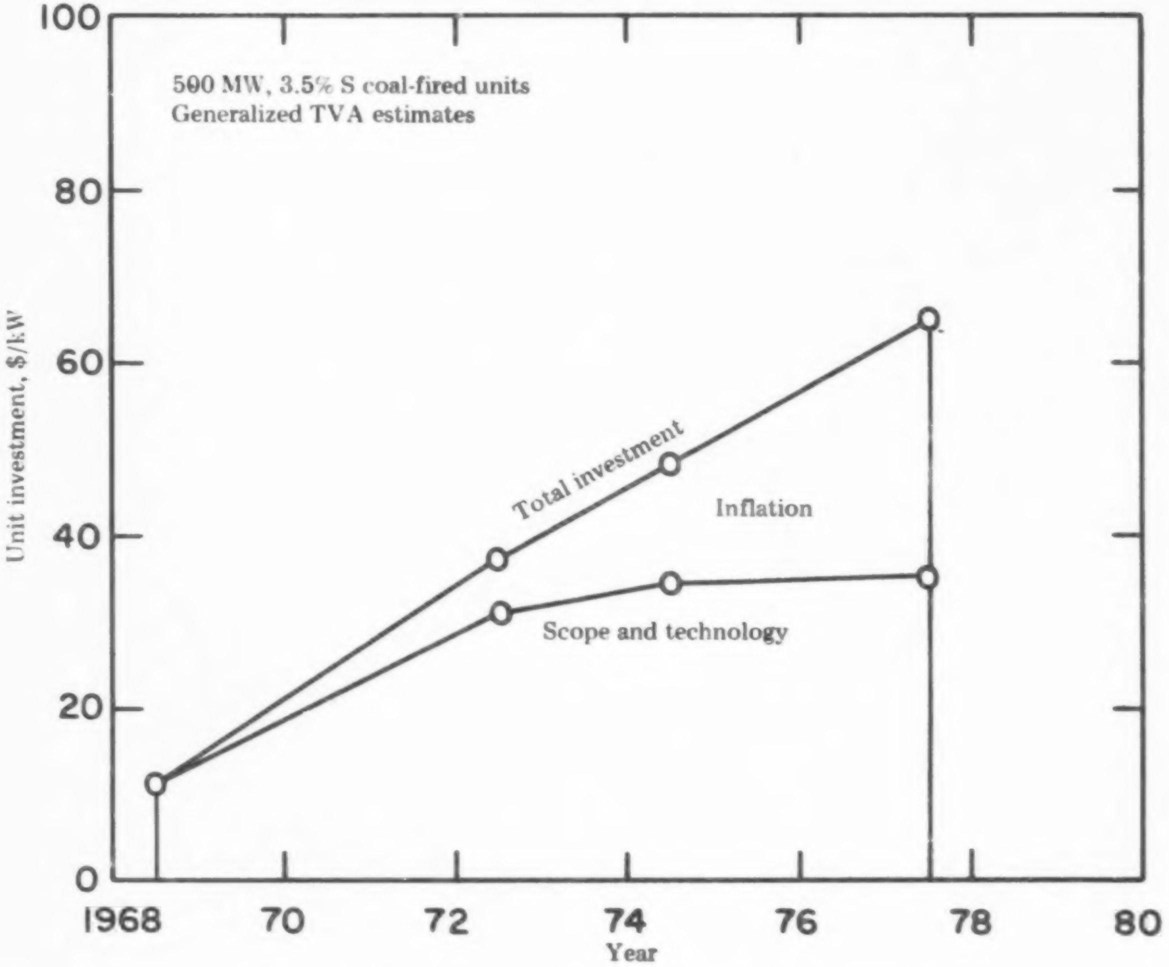


Figure 1. Limestone slurry scrubbing process - change in investment cost during past decade.

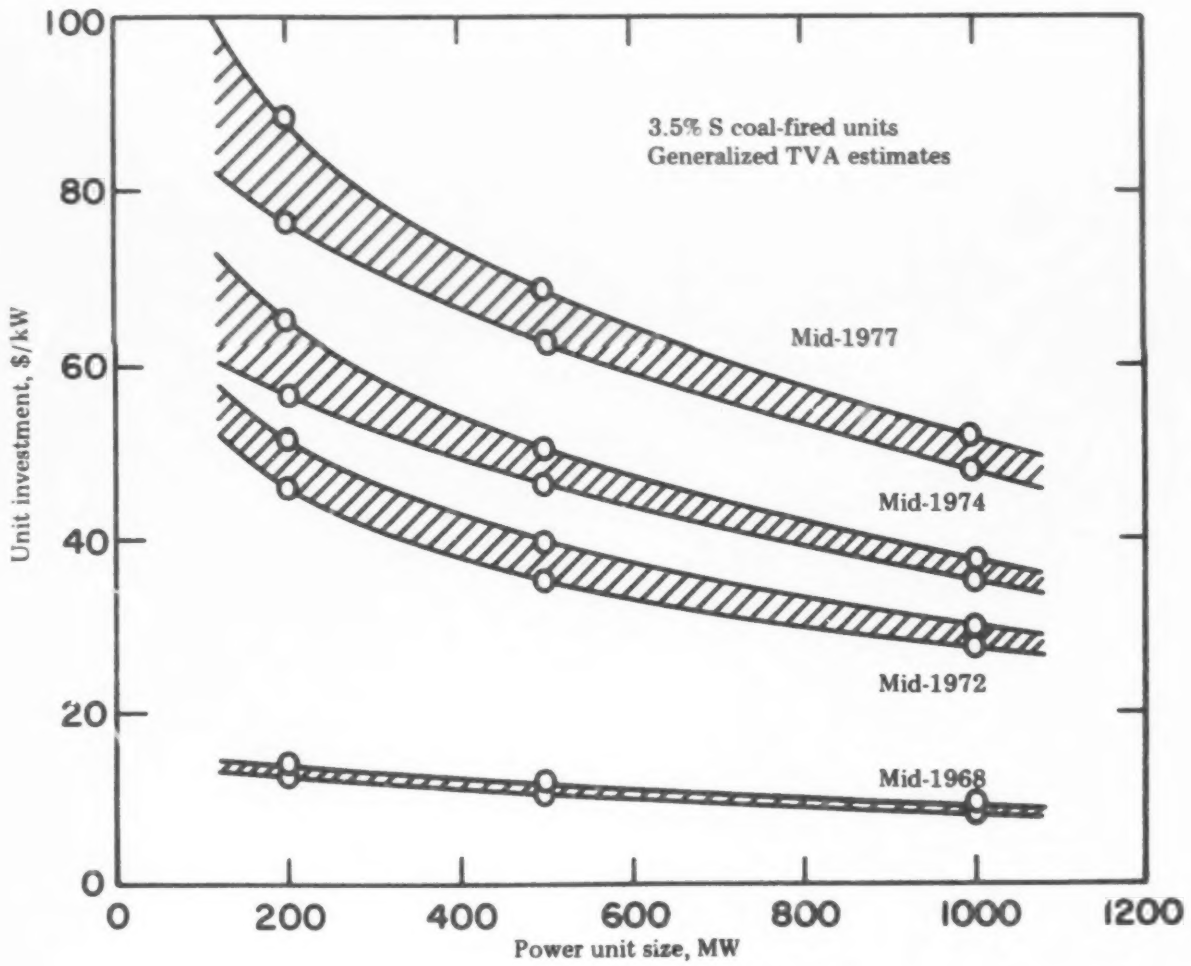


Figure 2. Limestone slurry scrubbing process - range of projected investment estimates.

UPDATE OF THE PAPER "COST COMPARISONS OF FLUE GAS DESULFURIZATION SYSTEMS"

Since 1967 the TVA Office of Agricultural and Chemical Development has been preparing for EPA or its predecessors cost estimates of various FGD processes. Many of these have been published or presented previously. Probably the best known set of EPA-TVA FGD process cost estimates was presented in a paper entitled "Cost Comparisons of Flue Gas Desulfurization Systems" at the EPA symposium in Atlanta, November 1974. Five processes were evaluated in a single comparative study--limestone slurry and lime slurry scrubbing; magnesia slurry - regeneration to 98% sulfuric acid; sodium scrubbing - regeneration to sulfur; and catalytic oxidation to 80% sulfuric acid. The complete data from which the paper was prepared were later published in January 1975 in an EPA report entitled "Detailed Cost Estimates for Advanced Effluent Desulfurization Processes." Because the project time frame for these estimates was mid-1972 to mid-1975 and capital and revenue requirements were subjected to tremendous escalation near the end of this period, TVA has updated these costs to reflect a mid-1975 to mid-1978 construction project. The revised investment costs are shown in Table 1 and the annual revenue requirements are shown in Table 2. (Revenue requirements include all operating costs, such as raw materials, labor, utilities, maintenance, overheads, capital charges, and taxes; assumes startup in mid-1978.) The basis and data used in updating these values are given in Table 3.

It should be noted that the revenue requirements shown in Table 2 do not reflect byproduct credits or sludge fixation costs. With recent increases in natural gas prices affecting the Frasch sulfur mining costs, we have seen sulfur list prices jump 85% (from \$35 to \$65/long ton) in 1975 with subsequent effect on sulfuric acid list prices (from \$30 to \$45/ton, 100% acid). Although these markets are now temporarily soft and spot prices reflect discounting, on a long-term basis, elemental sulfur prices can be expected to continue to climb with acid prices to follow. Introduction of large quantities of abatement byproducts would certainly change the picture, but if smaller quantities are introduced over a reasonable period of time (at the same rate as the total market growth), market disruption could be held to a minimum. Based on recent work (paper entitled "Potential Utilization of Controlled SO_x Emissions from Power Plants in Eastern United States" by J. I. Bucy, et al., presented at this meeting) TVA has completed for EPA on byproduct marketing, it would not be unreasonable to expect a long-term net revenue of \$45/long ton of sulfur or \$25/ton of 100% sulfuric acid.

Recent EPA policy statements now seem to be encouraging sludge fixation or similar long-term sludge disposal practices. If such costs must be added to those projected in Table 2 for limestone and lime scrubbing, the competitive position of these processes with respect to regenerative processes could change drastically. As illustrated in Table 4, the effect of byproduct credits for the regenerable processes and sludge fixation costs for the nonregenerable systems are significant. Since 90% or more of the FGD systems operating or under construction are the nonregenerative type, the utilities obviously need to take a closer look at the regenerative systems.

Table 1. SUMMARY OF TOTAL CAPITAL INVESTMENT REQUIREMENTS

FIVE LEADING FLUE GAS DESULFURIZATION PROCESSES

1975-78 COST BASIS

Case	Years life	Total capital investment[a,b,c]									
		Limestone		Lime		Magnesia		Sodium		Cat-Ox	
		M $	$/kW	M $	$/kW	M $	$/kW	M $	$/kW	M $	$/kW
Coal-fired power unit											
90% SO_2 removal; onsite solids disposal[e]											
200 MW, new, 3.5% S	30	17,671	88.4	15,978	79.9	19,119	95.6	22,391	112.0	25,171	125.9
200 MW, existing, 3.5% S	20	15,282	76.4	17,527	87.6	19,210	96.1	23,597	118.0	22,914	114.6
500 MW, existing, 3.5% S	25	31,113	62.2	35,001	70.0	34,940	69.9	43,351	86.7	48,772	97.5
500 MW, new, 2.0% S	30	30,734	61.5	27,545	55.1	31,057	62.1	36,802	73.6	54,494	109.0
500 MW, new, 3.5% S	30	34,185	68.4	30,558	61.1	35,827	71.7	42,387	84.8	54,791	109.6
500 MW, new, 5.0% S	30	37,114	74.2	33,143	66.3	39,941	79.9	47,210	94.4	55,047	110.1
1,000 MW, existing, 3.5% S	25	47,529	47.5	51,620	51.6	52,293	52.3	67,038	67.0	80,497	80.5
1,000 MW, new, 3.5% S	30	51,362	51.4	44,891	44.9	53,013	53.0	64,203	64.2	88,893	88.9
80% SO_2 removal; onsite solids disposal[e]											
500 MW, new, 3.5% S	30	32,955	65.9	29,428	58.9	34,659	69.3	40,406	80.8	-	-
90% SO_2 removal; onsite solids disposal[e] (existing unit without existing particulate collection facilities)											
500 MW, existing, 3.5% S	25	40,314	80.6	35,119	70.2	43,145	86.3	52,419	104.8	56,585	113.2
Oil-fired power unit											
90% SO_2 removal; onsite solids disposal[e]											
200 MW, new, 2.5% S	30	11,278	56.4	12,892	64.5	12,002	60.0	14,335	71.7	16,665	83.3
500 MW, new, 1.0% S	30	17,778	35.6	21,739	43.5	17,224	34.4	21,001	42.0	35,552	71.1
500 MW, new, 2.5% S	30	21,189	42.4	24,778	49.6	21,871	43.7	26,500	53.0	35,814	71.6
500 MW, new, 4.0% S	30	23,889	47.8	27,149	54.3	25,588	51.2	30,925	61.9	36,014	72.0
500 MW, existing, 2.5% S	25	25,215	50.4	29,329	58.7	27,309	54.6	33,618	67.2	42,366	84.7
1,000 MW, new, 2.5% S	30	32,866	32.9	36,193	36.2	32,377	32.4	40,590	40.6	58,122	58.1

---

[a] Midwest plant location represents project beginning mid-1975, ending mid-1978. Average cost basis for scaling, mid-1977. Minimum in process storage; only pumps are spared. Investment requirements for disposal of flyash excluded. Construction labor shortages with accompanying overtime pay incentive not considered.

[b] These investment costs depend heavily on project definition.

[c] Working capital has been included in the total capital investment.

[d] All Cat-Ox installations require particulate removal to 0.005 gr/acf prior to entering converter. Because existing units are assumed to already meet EPA standards (0.1 lb particulate/MM Btu of heat input), only incremental additional precipitator is required.

[e] Sludge disposal pond with clay liner.

Table 2. SUMMARY OF TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS

FIVE LEADING FLUE GAS DESULFURIZATION PROCESSES

1978 COST BASIS

Case	Years life	Total average annual revenue requirements[a,b]									
		Limestone		Lime		Magnesia		Sodium		Cat-Ox	
		M $	Mills/kWh	M $	Mills/kWh	M $	Mills/kWh	M $	Mills/kWh	M $	Mills/kWh
Coal-fired power unit											
90% SO_2 removal; onsite solids disposal											
200 MW, new, 3.5% S	30	5,883	4.20	6,362	4.54	7,036	5.03	9,238	6.60	6,000	4.29
200 MW, existing, 3.5% S	20	5,686	4.06	7,150	5.11	7,257	5.18	10,868	7.76	8,263	5.90
500 MW, existing, 3.5% S	25	11,854	3.39	14,528	4.15	14,052	4.01	22,189	6.34	17,765	5.08
500 MW, new, 2.0% S	30	10,625	3.04	11,145	3.18	11,572	3.31	14,549	4.16	12,681	3.62
500 MW, new, 3.5% S	30	11,937	3.41	12,758	3.65	14,082	4.02	18,782	5.37	12,766	3.65
500 MW, new, 5.0% S	30	13,105	3.74	14,377	4.11	16,448	4.70	22,858	6.53	12,844	3.67
1,000 MW, existing,3.5% S	25	19,711	2.82	23,819	3.40	23,169	3.31	38,813	5.54	31,138	4.45
1,000 MW, new, 3.5% S	30	19,163	2.74	20,570	2.94	22,789	3.26	31,186	4.46	20,534	2.93
80% SO_2 removal; onsite solids disposal											
500 MW, new, 3.5% S	30	11,319	3.23	12,206	3.49	13,406	3.83	17,425	4.98	-	-
90% SO_2 removal; onsite solids disposal (existing unit without existing particulate collection facilities)											
500 MW, existing, 3.5% S	25	14,376	4.11	14,826	4.24	16,639	4.75	24,995	7.14	19,480	5.57
Oil-fired power unit											
90% SO_2 removal; onsite solids disposal											
200 MW, new, 2.5% S	30	4,236	3.02	5,123	3.66	4,615	3.30	6,463	4.62	3,902	2.79
500 MW, new, 1.0% S	30	7,377	2.11	8,920	2.55	6,848	1.96	8,938	2.55	8,260	2.36
500 MW, new, 2.5% S	30	8,548	2.44	10,585	3.02	9,019	2.58	13,075	3.74	8,181	2.34
500 MW, new, 4.0% S	30	9,566	2.73	11,886	3.40	10,967	3.13	17,042	4.87	8,033	2.30
500 MW, existing, 2.5% S	25	9,979	2.85	12,140	3.47	10,683	3.05	15,322	4.38	15,998	4.57
1,000 MW, new, 2.5% S	30	13,317	1.90	17,165	2.45	14,734	2.10	22,223	3.17	13,125	1.87

---

[a] Power unit on-stream time, 7,000 hr/yr. Midwest plant location, 1978 revenue requirements. Investment and revenue requirements for disposal of flyash excluded.

[b] These revenue requirements reflect capital investments shown in Table 1 (updated); byproduct credit and sludge fixation costs excluded.

Table 3. DATA USED IN REVISION OF TABLES 11 AND 22 OF THE TVA PAPER "COST COMPARISONS OF FLUE GAS DESULFURIZATION SYSTEMS"

Data for Revision of Capital Investment

Investment basis: 500-MW, new coal-fired unit burning coal with 12% ash, 30-yr life, 127,500 hr of operation, only pumps spared, no bypass ducts, no overtime, experienced design and construction team, 3-yr project, 1975-78; 1977 costs for scale, includes both flyash and SO_2 removal, excludes flyash disposal, reheat to 175°F.

	Mid-1974	Mid-1977
Material index for direct costs	153.8	220.4
Labor index for direct costs	177.9	197.0

Data for Revision of Annual Revenue Requirements

Raw materials	Cost basis	1975	1978
Limestone	$/ton	4.00	6.00
Lime (varying with quantity)	$/ton	18.50-24.00	33.00-42.00
Magnesium oxide	$/ton	155.00	215.00
Coke	$/ton	15.00	28.00
Soda ash	$/ton	52.00	78.00
Catalyst (MgO, Cat-Ox)	$/liter	1.65	2.20
Catalyst (sodium)	$/yr	-	1.5 x 1975 cost
Antioxidant	$/lb	2.00	2.75
Conversion costs			
Labor: Operating, supervision	$/man-hr	8.00	10.00
Analyses	$/man-hr	12.00	15.00
Utilities			
Process water (vs. quantity)	$/M gal	0.02-0.08	0.03-0.11
Fuel oil, No. 6	$/gal	0.23	0.35
Fuel oil, No. 2	$/gal	0.30	0.42
Natural gas	$/Mcf	1.00	2.00
Heat credit, coal	$/MM Btu	0.60	1.15 ($22.80/ton coal)
Heat credit, oil	$/MM Btu	1.60	2.40

	Year	Cost basis	200 MW	500 MW	1,000 MW
Steam					
Coal	1975	$/M lb	0.80	0.70	0.60
Coal	1978	$/M lb	1.50	1.40	1.30
Oil	1975	$/M lb	1.50	1.40	1.30
Oil	1978	$/M lb	2.75	2.65	2.55
Electricity					
Coal	1975	$/kWh	0.011	0.010	0.009
Coal	1978	$/kWh	0.028	0.027	0.026
Oil	1975	$/kWh	0.019	0.018	0.017
Oil	1978	$/kWh	0.041	0.040	0.039

Table 4. REVENUE REQUIREMENT COMPARISONS (1978) OF FGD SYSTEMS INCLUDING BYPRODUCT CREDITS AND SLUDGE FIXATION COSTS

Process	Tons byproduct	Annual revenue requirements, mills/kWh As shown in Table 2[a]	 Includes byproduct credit[b]	 With sludge fixation[c]
Limestone	206,000[d]	3.41	-	4.01
Lime	175,000[d]	3.65	-	4.15
Magnesia	110,400[e]	4.02	3.23	-
Cat-Ox	137,400[f]	3.65	3.04	-
Sodium	32,700[g]	5.37	4.99	-

[a] New 500-MW, coal-fired unit, 3.5% S in coal, 7,000 hr/yr.

[b] 100% sulfuric acid at \$25/ton, 80% acid at \$16/ton, sulfur at \$45/long ton (\$40.18/ton).

[c] Assuming sludge fixation service fee by contract treating sludge in the utility's pond at \$10/ton of 100% solids.

[d] 100% solids.

[e] 100% H_2SO_4.

[f] 80% H_2SO_4.

[g] Elemental sulfur.

These updated economics unfortunately do not reflect any recent changes in process technology which, of course, should be taken into account. In addition to sludge fixation, numerous other developments in lime and limestone technology have taken place recently including (1) spray towers in place of mobile-bed scrubbing devices, (2) series hold tanks, magnesium addition and chloride ion control, pH and stoichiometry adjustments for increasing limestone utilization, (3) mist eliminator wash cycles and hot air injection or recycle reheat schemes to increase operating reliability, and (4) sludge oxidation for more effective disposal. Furthermore, in spite of the increased costs, some utilities have been leaning toward scrubber redundancy and equipment sparing as means for improving system reliability.

Although the cost effects of these changes have not been explored in updating the 1974 EPA-TVA cost study results, some of them have been examined as part of a separate study which will be presented as part of this paper.

For the three regenerable processes, there have not been as many noteworthy developments as with lime-limestone; however, some new information is available. For instance, data from the Boston Edison Company and Potomac Electric Power Company (PEPCo) magnesia scrubbing demonstrations indicate the process chemistry is workable as predicted, but some improvements are needed in liquid-solids separation, drying, calcining and materials handling equipment, and rubber-lined piping is definitely required. Since the piping in the EPA-TVA estimates was rubber lined, the expected cost effects of the above improvements would probably be limited to less than 10% over the 1975-78 projected costs.

Changes in the Wellman-Lord - Allied process primarily center around SO_2 regeneration where multiple-effect evaporation can be used to reduce energy requirements at the expense of additional capital. Probably, a savings of 5-6% in operating cost could be obtained with 2-4% increase in investment. It is also anticipated that some measures may be taken to reduce or eliminate the byproduct Na_2SO_4 generated by the process; however, such process alterations are yet under study and remain to be firmly applied.

Although the Wood River Cat-Ox demonstration has been plagued with a variety of problems, many unrelated to the process itself, we are not aware of any improvements or changes which would significantly alter the economics as presented.

Once these and other regeneration process demonstrations are carried to successful completion, there no doubt will be greater acceptance by the utilities. The regeneration processes badly need a success story, without any qualifications, to improve their image.

EXPANDED EPA-TVA PROCESS ECONOMICS PROGRAM

In the months since the report "Detailed Cost Estimates for Advanced Effluent Desulfurization Processes" was completed, TVA, under the sponsorship of EPA, has begun a larger, expanded program in evaluating FGD process economics which we hope will produce additional valuable results. The phases of study are as follows:

Economics of lime-limestone scrubbing based on results of the Shawnee test program

Byproduct marketing

Advanced process evaluations

Sludge disposal design and cost study

Energy optimization studies

Economics of Lime-Limestone Scrubbing Based on Results of the Shawnee Test Program

Joint Bechtel/TVA Computer Study. Within the scope of the Shawnee lime-limestone wet scrubbing test program, EPA is funding a joint study by Bechtel Corporation and TVA to develop a computer program which will predict investment and revenue requirements for various designs and/or operating conditions. This project has been under way since early 1975. The program can be used as a preliminary design and cost guide in choosing a lime-limestone scrubbing system for a particular application (e.g. limestone vs. lime scrubbing utilizing TCA scrubbers). It can also be used to determine the effect of a change of independent variables on the cost of a particular system (e.g. 8 ft/sec vs. 12.5 ft/sec scrubber gas velocity). The model should not be counted on, however, to compute the economics of a given system to a high degree of accuracy.

Bechtel's major responsibility in the study is to provide design equations for relating equipment size or capacity to the input process design variables. TVA is responsible for (1) the writing of the overall computer program, (2) the generation of tables and/or equations relating equipment cost to size or capacity, and (3) the estimation of total investment and revenue requirements, based upon the equipment costs and other economic inputs.

The overall program is not yet complete; however, Bechtel has completed the initial effort for specifying material flow rates and equipment sizes, and TVA has developed the methods for costing the equipment. The initial version of the overall program is expected to be complete by mid-1976. Upon completion of the computer program, a user will be capable of comparing the effect of various design conditions on relative economics of lime-limestone systems for coal-fired power units ranging from 100 to 1300 MW. As a further introduction to the program, the major input requirements for the joint Bechtel/TVA computer program are listed below.

1. Power unit size, MW
2. Boiler heat rate, Btu/kWh

3a. Coal analysis, wt % as fired

    Carbon
    Hydrogen
    Oxygen
    Nitrogen
    Sulfur
    Chloride
    Ash
    Moisture

3b. Heating value of coal, Btu/lb

4. Percent of sulfur emitted as SO_2 or SO_3 overhead
5. Percent of ash emitted overhead
6. Excess air to boiler, %
7. New versus retrofit installation
8. Limestone or lime chemical and physical properties
9. TCA scrubber operating parameters

    Number of scrubbers
    Number of beds
    Scrubber gas velocity
    Liquid to gas ratio
    SO_2 removal, %
    Stoichiometry

10. Redundancy desired

    Number of spare scrubbing trains

11. Mist eliminator system options

    With or without wash tray

12. Effluent hold tank definition

    One, two, or three stirred tanks (series)
    Slurry residence time, min (total)

13. Reheat (in-line steam reheater)

    Outlet flue gas temperature, °F

14. Electrostatic precipitator or mechanical collector

    Ash removal efficiency upstream of SO_2 scrubbers

15. Sludge disposal method

    Onsite ponding: Distance between disposal site and scrubber system
    Weight percent solids in settled sludge

    Fixation: Method or costs

16. Power unit load factors
17. Remaining life of plant
18. Economic factors

    Construction cost indices
    Raw material and utility unit costs

The major outputs of the joint Bechtel/TVA computer program are listed below.

1. Conceptual design information

    Material balance
    Number of trains
    Equipment size, general specifications, and operating conditions

2. Capital investment

    Equipment costs
    Investment breakdown by area (feed preparation, scrubbing, disposal)
    Investment breakdown by item (equipment, piping, ducts, structure, electrical, instrumentation, etc.)

3. Annual revenue requirement

    Direct costs: Raw materials
    Labor
    Utilities
    Maintenance

    Indirect costs: Capital charges
    Overheads

    Total annual revenue requirement

4. Lifetime revenue requirement

    Total
    Average unit
    Levelized unit

EPA-TVA Limestone Process Economic Sensitivity Study. In April 1975 TVA completed a sensitivity study for EPA showing the effect of several selected operating variables on the costs of limestone scrubbing. As an example of the potential use of the computer program upon completion, the results of selected design options considered in the April 1975 sensitivity study have been revised manually to reflect more accurately the equipment definition for the computer study. The updated results are shown in Table 5 as ratios to total investment and total revenue requirements of the base case. In the computer study, results will be calculated to give a total project cost rather than as a ratio to a base case.

Table 5. LIMESTONE SLURRY PROCESS

PROJECTED INVESTMENT AND REVENUE REQUIREMENT RATIOS[a]

Case	Identification	Total investment ratio	Total annual revenue requirement ratio
1	Base case	1.000	1.000
2	12.5 ft/sec gas velocity	0.915	0.935
3	12.5 ft/sec gas velocity with wash trays	0.931	0.975
4	MgO addition	0.911	0.974
5a	Hypalon-lined pond	1.125	1.070
5b	Pond without clay lining	0.981	0.989
6a	Sludge fixation service fee - \$5/wet ton (50% solids)	0.839[b]	1.099
6b	Sludge fixation service fee - \$10/wet ton (50% solids)	0.839[b]	1.287
7	Slurry oxidation - low pH operation with series hold tanks	1.033	1.140
8	Slurry oxidation - MgO addition	1.030	1.177
9	Benzoic acid addition	0.990	1.008
10	Series hold tanks	0.997	0.990
11	Low pH operation	0.990	0.994
12	Low pH operation with series hold tank	1.006	1.001
13a	Hot air injection reheat - 100°F	1.080	1.209
13b	Hot air injection reheat - 50°F	1.022	0.989
14a	Flue gas recirculation reheat - 100°F	1.145	1.100
14b	Flue gas recirculation reheat - 50°F	1.037	0.958
15	In-line reheat - 50°F	0.984	0.920
16	In-line reheat - 100°F - entrainment level - 0.67%	1.001	1.032

---

[a] 500-MW new coal-fired power unit, 3.5% S in fuel, 7,000 hr/yr.
85% SO_2 removal, mid-1977 investment, mid-1978 revenue requirements.

[b] Assumes service contractor provides investment for sludge transportation, fixation, and disposal. Includes only cake filtration and loading facilities - pond excluded.

Base Scrubbing System Incorporated in the Joint Bechtel/TVA Computer Program and the Economic Sensitivity Study. The base system reflected in the computer program and sensitivity study utilizes a forced-draft, dry fan system upstream of the SO_2 scrubbers. Costs include a low-efficiency mechanical collector upstream of the fans to insure removal of the large ash particles in the gas thereby preventing erosion of the fan blades. A plenum is used for separation of the power plant ducts from the scrubber ducts, allowing the number of scrubber trains to be a variable, independent of the number of power plant ducts. Although separate fans are provided on each side of the plenum to insure satisfactory gas distribution, only the scrubber fan costs are included within the estimates. The SO_2 scrubber is designed with a presaturator compartment for partial humidification and cooling of the gas upstream of the first bed. The number of scrubber beds and depth of spheres per bed are inputs.

An in-line steam reheater is located downstream of the scrubber in the vertical section of duct. The reheater is designed with Inconel 625 tubes for reheating the gas to 150°F. All reheater tubes in contact with the gas above 150°F are Corten. The gas is reheated to a final temperature of 225°F.

With the exception of the "fixed sludge" variations, each case is designed and sized for onsite pond disposal of wet collected flyash in addition to the SO_2 byproducts. However, costs for transporting and disposing the ash removed by the mechanical collectors are excluded. The distance between the scrubbers and the disposal pond for each case is 1 mi. The concentration of solids in the settled sludge is assumed to be 40% by weight for all cases except those designed for high oxidation of sulfites to sulfates. The concentration of settled solids in the pond for the high oxidation cases is assumed to be 60% by weight. Disposal lines (spares included) to the pond are rubber-lined carbon steel. Supernate return lines are carbon steel. Ponds are sized for a total of 127,500 hr of operation over the life of the plant, corresponding to an average capacity factor of 48.5% over a 30-yr plant life.

A detailed listing of the cases and premises for these updated costs and the initial computer program is shown below.

Definition of Cases for Updated TVA Economic Sensitivity Analyses.

Base Case

Case 1: Limestone additive
Megawatts = 500 (4 modules)
Sulfur in coal = 3.3 wt %; ash in coal = 12 wt %
TCA with 3 stages (4 grids), 5 in. spheres/stage
Scrubber pressure drop = 4.2 in. H_2O (excluding mist eliminator)
Chevron mist eliminator pressure drop = 0.2 in. H_2O
L/G = 75 gal/Mcf
Liquor rate = 36 gpm/ft^2
Gas velocity = 8 ft/sec
Limestone utilization = 65%
SO_2 removal = 85%
Particulate removal = 99.5%
Solids in recirculated slurry = 15 wt %
Scrubber inlet liquor pH = 5.8
Effluent residence time (single tank) = 12 min
Pond (clay-lined) for sludge disposal
In-line steam (500 psig pressure) reheater, 100°F reheat
FD fan, with partial coarse ash cleanup, 33% efficient mechanical collector
Scrubber inlet particulate loading = 2-3 gr/scf

Velocity Effect

Case 2: Same as Case 1, except:
Gas velocity = 12.5 ft/sec
Liquor rate = 33 gpm/ft^2
L/G = 44 gal/Mcf
Scrubber pressure drop = 7.2 in. H_2O (excluding mist eliminator)
Chevron mist eliminator pressure drop = 0.3 in. H_2O

Case 3: Same as Case 1, except:
Gas velocity = 12.5 ft/sec
Liquor rate = 33 gpm/ft^2
L/G = 44 gal/Mcf
•Wash tray plus chevron mist eliminator
Scrubber pressure drop = 7.2 in. H_2O (excluding mist eliminator and wash tray)
Chevron mist eliminator pressure drop = 0.3 in. H_2O
Wash tray pressure drop = 2.7 in. H_2O

MgO Addition Effect

Case 4: Same as Case 1, except:
Liquor magnesium ion concentration = 10,000 ppm
Effluent residence time = 5 min
L/G = 40 gal/Mcf
Liquor rate = 19.2 gpm/ft^2
Scrubber pressure drop = 3.1 in. H_2O (excluding mist eliminator)

Alkali Utilization/Sludge Disposal Effects

Case 5a: Same as Case 1, except Hypalon-lined pond

Case 5b: Same as Case 1, except pond without clay lining

Case 6a: Same as Case 1, except no pond; sludge fixed and disposed under service contract of $5/wet ton (50% solids)

Case 6b: Same as Case 6a, except no pond; sludge fixed and disposed under service contract of $10/wet ton (50% solids)

Case 7: Same as Case 1, except:
12 in. spheres/stage
Scrubber pressure drop = 7.2 in. H_2O (excluding mist eliminator)
Limestone utilization = 90%
Scrubber inlet liquor pH = 5.2
Three series hold tanks (4 min each)
Complete oxidation of slurry bleed stream (pH ~4.5) from the first hold tank, using air at 5 atm and stoichiometry of 5 moles O_2/mole SO_2 absorbed (similar to JECCO mode).

Case 8: Same as Case 4, except complete oxidation of slurry bleed stream (from the single hold tank) using air at 3 atm and stoichiometry of 5 moles O_2/mole SO_2 absorbed (similar to Kellogg mode). Use 1 hr residence time in the oxidizer.

Case 9: Same as Case 1, except:
Organic additive
Limestone utilization = 80%

Case 10: Same as Case 1, except:
Three series hold tanks (4 min each)
Limestone utilization = 75%

Case 11: Same as Case 1, except:
Scrubber inlet liquor pH = 5.2
Limestone utilization = 80%
12 in. spheres/stage
Scrubber pressure drop = 7.2 in. H_2O (excluding mist eliminator)

Case 12: Same as Case 7, except no oxidation of slurry bleed stream.

Reheater Effect

Case 13a: Same as Case 1, except:
External steam reheater with air - 100°F

Case 13b: Same as Case 13a, except - 50°F

Case 14a: Same as Case 1, except:
External steam reheater with flue gas recycle - 100°F

Case 14b: Same as Case 14a, except - 50°F

Case 15: Same as Case 1, except - 50°F

Case 16: Same as Case 1, except entrainment level - 0.67% of scrubber outlet flue gas rate

Byproduct Marketing

The second most advanced project in the overall EPA-TVA FGD economics program is an intensive study of marketing byproducts from FGD systems. Byproducts from both nonregenerable and regenerable processes are being evaluated for potential market volume, location, and price.

The initial efforts in this area were undertaken in early 1973 when a hypothetical study was made of marketing sulfuric acid which TVA could conceivably produce from its coal-fired power plants. Using a computerized transportation model a report entitled "Marketing H_2SO_4 from SO_2 Abatement Sources--The TVA Hypothesis" was published in December 1973 indicating that approximately 2 million tons per year of abatement acid could be produced. Under the market conditions prevailing at that time, a net sales revenue of $5-9/ton was indicated. With recent price increases, no doubt higher revenues could now be expected.

In June 1974 TVA prepared a second byproduct marketing study for EPA entitled "Preliminary Feasibility Study of Calcium-Sulfur Sludge Utilization in the Wallboard Industry." Although limited in scope and very brief, this study indicated the possible savings to be derived from marketing sludge as gypsum.

With these two projects as pilots, agreement was reached in 1974 to expand the sulfuric acid analyses to cover both sulfuric acid and elemental sulfur for all the utilities in the United States and to carry out a more comprehensive study of gypsum markets. The computer program used for the TVA hypothetical assessment was to be modified and expanded considerably for this effort.

During the early phases of the expanded acid study, it became apparent that a sophisticated, useful computer program applicable to any byproduct was being developed and plans for a further enlargement of the byproduct market investigation series were initiated. Current plans now provide for expansion of the studies to include ammonium sulfate and relocation of the phosphate fertilizer industry to the Midwest plus a systems study to determine which byproduct would be best for candidate coal-fired power units in the United States. When completed, the systems program will be available to users on a national time-sharing network. The initial phase of this work is covered in a separate paper which will be presented later in the symposium.

As part of this phase of work considerable cost estimating has been required to derive competitive economics for commercial contact sulfuric acid manufacture including tail gas cleanup, storage terminal costs for acid and sulfur, Frasch sulfur mining costs, smelter SO_2 control costs, and acid neutralization costs. All of these are subprogrammed in the overall systems analysis.

Toward evaluating gypsum production and marketing economics, TVA has begun an appraisal of processes capable of producing gypsum from sludge or scrubber effluent. Those systems to be studied will include the Chiyoda dilute acid system, the Mitsubishi - JECCO oxidation process, and the M. W. Kellogg oxidation system. This conceptual design and cost study will be used to compare the economics of utilization for wallboard or other uses versus disposal as waste.

Advanced Process Evaluations

A third phase of work now under way provides for design and economic evaluations of the latest advanced processes being considered for SO_2 control. Both double-alkali processes which produce sludge and regenerable processes producing sulfur are already under study.

In support of the EPA demonstration programs, preliminary, rough assessments (new 500-MW, coal-fired units, 3.5% S in coal) were made of the Atomics International's aqueous carbonate, UOP - Shell's copper oxide, Air Products - Catalytic's ammonia/IFP, and Consol's potassium scrubbing processes. These processes all produce elemental sulfur. Preliminary results (yet to be confirmed by detailed study) indicate a wide range of investment ($87 to $160/kW for 1975-78 project) and total annual revenue requirements (3.7 to 7.5 mills/kWh for 1978 startup). If $45/long ton of sulfur is taken as byproduct credit, the net revenue requirements range from 3.3 to 7.1 mills/kWh.

In addition to the regenerable processes, three double-alkali systems--Envirotech, CEA - ADL, and FMC--have been studied. The total investment (including sludge pond) and revenue requirements of the three systems are relatively close, averaging about $71/kW investment and 3.6 mills/kWh annual revenue requirement. Additional refinement of vendor data will be necessary before true comparative costs can be derived.

Thus far, in all processes evaluated, capital charges (excluding maintenance) constitutes the largest fraction of annual revenue requirements ranging from about 40-43% for the double-alkali processes to about 45-60% for the regenerable processes. In the double-alkali processes, raw materials (primarily lime) require the second greatest expenditure with energy third. This is reversed with regenerable processes in that they require more energy (fuels, electricity, and steam) than raw materials. (This can be misleading since components such as naphtha, coke, coal, and fuel oil could easily be classified as raw materials or energy.) Operating labor cost for all the processes is less than 3% of total revenue requirements.

Plans now call for more intensive evaluation of the most promising systems. Detailed work is already under way on the Bureau of Mines' citrate process for which design, investment, and revenue requirement data will be developed along the lines of the earlier EPA-TVA studies. After review by the process developers for accuracy, reports will be published as with previous evaluations.

Sludge Disposal Design and Cost Study

Another area of work to be covered in the process economics program is a detailed conceptual design and cost study of sludge disposal processes. In the past the lack of detailed design and cost data for this area has reduced the value of our process economic assessment effort. Included will be cases covering several pond design and liner options, fixation processes such as IUCS, Dravo, and Chemfix, and various sludge dewatering techniques. In addition, accurate design and cost data will be developed for combined and separated flyash - scrubber sludge disposal systems. This work is just getting under way and the results will also be published.

Energy Optimization Studies

The final phase of economic activity covers a careful investigation of SO_2 removal process energy requirements and is directed toward reducing or minimizing process energy needs. An effort will be made to obtain data from actual scrubber installations to compare with theoretical needs. In addition, key unit operations will be analyzed toward finding alternate routes to lower energy consumption. Work on this project is to be started soon.

ECONOMICS FOR THE TVA OFFICE OF POWER

In addition to the work being performed for EPA, the OACD studies staff has continued to support the TVA Office of Power with specific energy-environmental-related process evaluations. In the area of FGD, evaluations are now under way on two promising Japanese double-alkali processes--Kureha and Dowa. The Kureha system uses a sodium acetate scrubbing solution to react with limestone whereas the Dowa process uses a soluble aluminum sulfate - oxide complex in water to absorb SO_2 for further reaction with limestone. Recent visitors to Japan have been impressed with these two processes which produce gypsum for sale or disposal by piling.

Results thus far indicate these two processes to be competitive with double-alkali systems under development in the United States. If sludge fixation or other extra processing measures (reduction of soluble salt losses, oxidation of sludge) are required, these processes will have strong advantages.

Also, during the past year, OACD has taken part in other Office of Power energy process evaluations such as solvent refined coal, fixed-bed coal gasification, NO_x removal studies, and fluidized-bed combustion. Some of these results have been reported previously. These activities have enabled us to make comparisons of the various power plant environmental control options with perspective and consistency.

As can be seen, TVA has been and continues to be involved in a number of studies on FGD system economics. In meeting these and future responsibilities, we are striving to keep our quality high so that representative process comparisons can continue to be made to guide the industry in making key decisions.

STATUS OF THE EPRI FLUE GAS DESULFURIZATION DEVELOPMENT PROGRAM

Lawrence W. Nannen and Kurt E. Yeager

Electric Power Research Institute
Palo Alto, California

ABSTRACT

This paper summarizes the present status of EPRI's program in flue gas desulfurization for both lime/limestone and advanced regenerable processes. Emphasis is given to problem areas in lime-limestone scrubbing since it offers the most widely accepted near-term solution to positive SO_2 emission control. Specific EPRI projects are discussed and general guidelines for reliable scrubber implementation are offered.

Preceding page blank

STATUS OF THE EPRI FLUE GAS DESULFURIZATION DEVELOPMENT PROGRAM

by

Lawrence W. Nannen
and
Kurt E. Yeager
Electric Power Research Institute
Palo Alto, California

The Electric Power Research Institute began operations in 1973 for the purpose of expanding electric energy research and development under the sponsorship of the nation's utility industry. EPRI is presently funded by over 500 member organizations, public, private, and cooperative. EPRI's overall goal is to develop a broad, coordinated, advanced technology program for improving electric power production, transmission, distribution, and utilization in an environmentally acceptable manner. The primary areas of EPRI's research are fossil fuel and advanced systems; nuclear power; power transmission and distribution; and energy systems, environment, and conservation.

Projects of high priority at EPRI include those which develop technological options to permit continued utilization of coal for power production in both new and retrofit applications within environmental constraints. A major current issue associated with all fossil fuel combustion, particularly coal, is the control of sulfur oxide emissions.

Current EPRI Objectives in Flue Gas Desulfurization

The present research and development projects dealing with SO_2 control within EPRI's Environmental Control and Combustion Program are oriented to providing electric utilities with reliable, cost-effective technology for removing sulfur oxides from coal-fired steam boilers. The goals of these projects are to ultimately develop a reliable commercial design basis for both throwaway scrubbing systems and regenerable FGD technology so that utilities can apply such systems if they are required.

LIME/LIMESTONE SCRUBBING

Since the technology for lime/limestone scrubbing is generally considered to be the most advanced for present commercial applications to utility boilers, and since the bulk of utility commitment is towards this technology (40,000 MW and $3 billion by 1980), the obvious near-term objective of EPRI is to assure that the technical issues are resolved to allow reliable, efficient operation of these systems. A large amount of operating experience now exists for lime/limestone scrubbing -- some good, some bad, and most poorly understood. Therefore, a major task is to collect the available data, sort out areas in which the technology needs improvement, and develop a methodology for designing these processes on a site-specific basis.

Table 1 lists the major problem areas continuing to face the utility industry in integrating lime/limestone scrubbing with power production.

Table 1

Major Problem Areas in Lime/Limestone Scrubbing

- Mist Elimination
- Reheat
- Materials of Construction
- Process Chemistry and System Design for Site Specific Applications
- Waste Handling and Disposal

As is the case with any complex process, one must maintain the awareness that each problem area is interrelated with another and must be addressed with the total system in mind.

Mist Eliminator

The function of a mist eliminator in any wet scrubbing system is to prevent the carryover of aqueous aerosol droplets downstream from the absorber to prevent plugging with slurry solids and corrosion of reheaters, scrubber I.D. fans, and flue gas duct work. The design of the mist eliminator system must consider the following variables:

Orientation. The demister may be placed horizontally in a vertical gas duct or vertically in a horizontal gas duct. It may also be inclined to assist drainage during washing. It appears that a vertical demister configuration (horizontal gas flow) offers advantages of superior drainage, greater washing flexibility, and better adaptability to closed-loop demister wash water recycle.

Flow Distribution and Velocity. Most mist eliminators used in commercial FGD scrubbing systems are of the zig-zag "Chevron" baffle design. Proper flow distribution and velocity of the wet gas through the demister is essential for proper operation and efficiency. The radial vane collector shows promise in improving gas distribution.

Washing. Reliable operation of mist eliminators in lime/limestone scrubbing systems is complicated by the normally high solids content of the scrubbing liquor. These liquids can adhere to the demister surface, forming a soft mud deposit or a hard chemical scale. A large amount of development work is presently underway to design a washing system to minimize solids buildup on the demister.

Tie-in with the Overall System. If the scrubber system is designed for closed-loop operation (no water discharge), the quantity of fresh water available to the system is limited to the water lost by saturation of the flue gas and water remaining with the waste solids after final disposal. Most experience to date clearly indicates that the demister wash water should be of the best quality, with the minimum amount of dissolved calcium, to avoid scaling. There is also a greater tendency for the demister to plug when limestone is used rather than lime as the absorbent. This is caused by the normally poorer utilization of the limestone which allows excess unreacted calcium to be entrained into the demister. The use of series hold tanks and low inlet slurry pH (between 5.0 and 5.5) can improve limestone utilization; however, some reduction in SO_2 removal efficiency may occur at low scrubbing pH.

Reheater

The function of a reheater is to warm the flue gas above its wet bulb temperature of 125°F after leaving the scrubber to a temperature normally between 175°F and 200°F. The warmer gas is then less likely to condense and corrode fans, duct work, and gas dampers. Reheat also increases plume rise and dispersion of the effluent to minimize ground-level concentrations of not only SO_2 but also particulates and nitrogen oxides.

Degree of Reheat Needed. There is some question as to the degree of reheat needed for site-specific meteorological conditions. There is no legal requirement for reheat in present standards.

In-Line Steam Tube Reheat. The most common design of reheat is a series of steam tubes inserted in the flue gas stream. Good reheater operation then depends largely on proper mist eliminator performance. However, corrosion is a major problem.

Indirect Reheat. A more costly, less efficient but very reliable design is to blend heated air with the flue gas to attain the desired temperature. This approach results in reduced maintenance costs.

Hot Gas Bypass. Use of hot gas bypass is dependent on the efficiency of the scrubber to remove a very high percentage of the SO_2 in the processed flue gas.

Fuel Combustion. The use of natural gas or fuel oil for direct combustion in the exit gas stream may be attractive in a few isolated cases, but is a poor option for the future due to the risk of fuel shortages.

Materials of Construction

Lime/limestone process slurries contain abrasive solids (limestone, fly ash, and gypsum) and corrosive dissolved species (chlorine, hydrogen, and concentrated salts). A major design problem is selection of reasonable cost materials of construction to minimize corrosion and erosion of scrubber components.

Liquor Loop. Pumps, valves, piping, and spray nozzles are all affected by process slurry conditions.

Flue Gas Handling. Fans, duct work, and the reheater must withstand the corrosive flue gas environment.

Absorber Internals. Scrubber internals such as the mist eliminator, stage partitions, trays, and packing must withstand both gas and liquor contact.

Process Chemistry and System Design

The previously discussed problem areas were primarily hardware issues. It is important to recognize that the poorest understanding of lime/limestone scrubbing is in the process chemistry and its effect on overall system design. We are still unable to confidently predict the chemical behavior of a scrubber for each potential utility site without a time-consuming, expensive pilot testing program at each site. Table 2 lists the important site-specific variables which affect scrubber system design.

The major scrubber chemistry issues requiring further study are:

Degree of Sulfite to Sulfate Oxidation. How is oxidation promoted or retarded in the scrubber? How is oxidation of the waste product achieved to allow better settling and dewatering?

Effect of Fly Ash Chemistry. Fly ash provides surface area for possible chemical reactions. Also, alkaline fly ashes from Western coals can be used beneficially in the scrubbing process.

Supersaturated vs. Unsaturated Gypsum Chemistry. We do not fully understand what conditions cause operation unsaturated with respect to calcium sulfate.

Table 2

Process Chemistry and System Design

Site-Specific Variables

Degree of SO_2 removal required

Inlet SO_2 concentration

Excess air

Degree of oxidation

Fly ash loading

Fly ash chemistry and solubility

Degree of closed-loop operation required

Dissolved solids in system

Tie-in with other plant waste water streams

Constraints on sludge disposal

Chemistry Affects System Design

Absorber type

Reaction tank

Liquid-to-gas ratio

Reactant type

Process control

Materials selection

Sludge composition and handling

Effect of Magnesium and Chlorine on scrubber chemistry.

Use of Lime vs. Limestone. Limestone is the most economical, energy efficient reagent but in many cases complicates the scrubber chemistry. More work is needed to understand how to increase limestone utilization while maintaining high SO_2 removals.

Use Cooling Tower Blowdown for Closed Loop Operation. The scrubber has the potential to act as a "garbage disposal" for the power plant. Water supplies can be conserved if scrubbers can feed off other plant waste streams.

Waste Handling and Disposal

Handling and disposal of lime/limestone scrubber sludge is a complex site-specific problem for which usable guidelines are not available. Scrubber wastes consist of three general types of material: fly ash, calcium sulfate/sulfite salts, and scrubbing liquor associated with the partially dewatered sludge. Each type of material has different physical and chemical characteristics. For example, trace elements are found almost exclusively in the fly ash, the calcium sulfite has very poor physical properties resulting in inadequate dewatering and structural stability features, and the liquor contains concentrated dissolved salts produced from the scrubbing process. Disposal and treatment methods have started to evolve without a clear understanding of each sludge component; the problems unique to one general component have been confused to be characteristic of the total sludge. Problems such as trace element leachability, sludge fixation, sludge dewatering, and beneficial use of scrubber wastes cannot properly be addressed until the characteristics of each sludge component are understood.

EPRI Task Areas for Throwaway SO_2 Scrubbing

The following discussion presents the methodology and organization of the R&D tasks in EPRI's throwaway SO_2 program.

1. Evaluate Utility Operating Experience (1974-75)

EPRI is funding the Battelle Stack Gas Coordination Center. Battelle has completed a state-of-the-art study for EPRI summarizing the commercial SO_2 experience in the U.S. and recommending further areas of research.

EPRI has also formed a utility working group to advise the EPA and TVA on the conduct of the Shawnee test program.

2. Unit Analyses of Scrubber and Disposal Hardware (1975-76)

Battelle is continuing their effort to evaluate scrubber problem areas with a series of "unit analyses" on mist elimination, reheat, and cost variation in scrubber bids.

A new contract is under negotiation with Southern California Edison and Stearns-Roger to study improved process control capabilities for scrubbers and the instrumentation available to measure control variables.

Proposed new projects to evaluate sludge handling techniques and to recommend pump, fan, and valve materials and design are planned.

EPRI is also supporting the Southern Services testing at Plant Sholz in Florida which is operating three 20 MW pilot plants: double alkali, dilute acid Chiyoda, and Foster Wheeler carbon adsorption.

3. Develop and Test Hardware and Chemistry (1976-77)

The Tennessee Valley Authority has been contracted by EPRI to study four tasks: a) horizontal 1 MW pilot scrubber for high sulfur coal, b) improved methods of reheat, c) corrosion/erosion materials testing, and d) characterization of scrubber sludge as a function of scrubber operating conditions.

A contract is under negotiation with Radian Corporation to demonstrate a novel concept in scrubber process control.

New projects include a study to characterize low sulfur/alkaline ash scrubbing chemistry for Western coals and an investigation into the oxidation chemistry of sulfite to sulfate.

4. Diagnostic Capability (1977-80)

It is planned that a large portion of EPRI's efforts in the future will be centered around diagnostic testing of specific fuels in several prototype scrubbers situated around the country. This service will benefit the power industry by providing a means to demonstrate overall scrubber operation and behavior before committing to a specific design. Diagnostic sites will be developed for both Eastern high sulfur coal and Western low sulfur coal.

Figure 1 describes the current EPRI SO_x Control Program and illustrates the interrelation and timing of specific projects for both lime/limestone scrubbing and regenerable FGD technology.

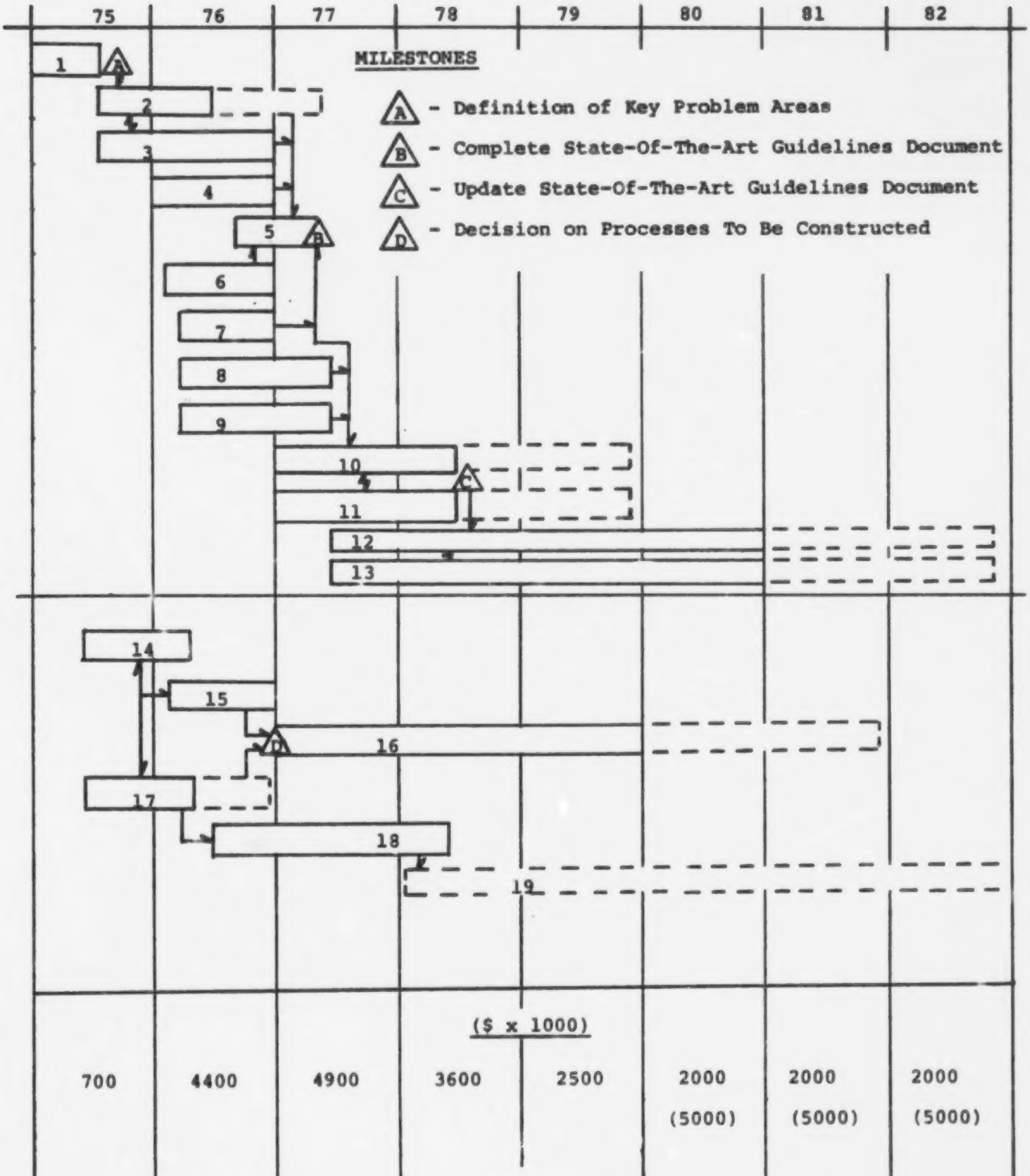
FIGURE 1

SO_x CONTROL PROGRAM

OBJECTIVE	PROJECT DESCRIPTION
Develop Reliable Design Basis for Lime/Limestone Scrubbing	1. Evaluate Utility Operating Experience
	2. Unit Studies-Demister, Reheat, Cost Variation
	3. 1 MW Tests-Reheat, Corrosion, Horizontal Scrubber
	4. Evaluate FGD Process Control Capability
	5. Design Guidelines for Lime/Limestone Scrubbing
	6. Low Sulfur/Alkaline Ash Scrubbing Characterization
	7. Evaluation of Sludge Dewatering Processes
	8. Sludge Composition & Leachability
	9. Sludge Fixation Chemistry Guidelines
	10. Chemistry Modifications To Improve Reliability & Cost
	11. Hardware Modification To Improve Reliability & Cost
	12. Diagnostic Support - Eastern Coal
	13. Diagnostic Support - Wstern Coal
Develop Viable Regenerable FGD Technology	14. Comparative Evaluation of Regenerable FGD Processes
	15. Comparative Design Of Advanced Processes
	16. Co-Sponsored Pilot (10-40MW) Process Construction & Operation
	17. 20 MW Pilot Test Of Chiyoda, Double Alkali, Bergbau Processes
	18. Modification To Existing Processes To Improve Reliability & Cost
	19. Test & Evaluation Of Commercial Scale (100MW+) Processes

FIGURE 1 (continued)

SO_x CONTROL PROGRAM



ADVANCED REGENERABLE PROCESSES

Recognizing the need to develop desulfurization processes which 1) do not create a solid waste disposal problem, 2) are not dependent on limestone raw materials, 3) have the possibility of being less sensitive to site-specific design variables, and 4) can produce a useable byproduct. EPRI is active in the evaluation of advanced regenerable processes. The objective of a current contract with Radian Corporation is to compare and evaluate a large number of advanced processes on technical grounds (material and energy balances) and to recommend further areas for process development.

EPRI's evaluation program is considering three broad areas of regenerable process development:

1) Modifications to commercially available regenerable FGD processes such as Wellman-Lord, magnesium oxide scrubbing, and CatOx to improve performance, reliability and cost.

2) Development of a regeneration capability for calcium sulfite/ sulfate throwaway FGD processes. In certain circumstances, particularly as a retrofit to lime/limestone scrubbing processes in the future, it may be possible to convert the sludge into sulfur or acid or to convert calcium based systems to magnesium oxide regenerable processes.

3) Development of new regenerable processes. A new EPRI project is planned for 1976 which will provide a competitive site-specific design evaluation for processes in all three of the above categories, with the possible selection of one or two new processes for pilot plant (10-40 MW) construction and demonstration with a host utility.

CONCLUSIONS

For the near term, the primary utility choice for SO_2 control appears to be throwaway scrubbing processes involving the use of alkaline reagents and the production of a calcium sulfite/sulfate waste product. Many utilities are reluctant to install these systems for three fundamental reasons:

1) Scrubbers are entirely non-profit for utilities and are expensive to install and operate. There is no guarantee that increased costs can be passed on to the customer in a timely fashion.

2) There is a questionable cost/benefit relationship and many utilities doubt the need to control SO_2 emissions to the degree many regulations require.

3) Early experiences with scrubbers were very poor and utilities have been frightened by their foreign nature to conventional power plant operation. Only recently have utilities recognized the fact that scrubbers are complex developmental chemical processes and operated them as such.

Considering these difficulties and the lack of positive incentive for the utility industry to purchase and operate chemical processes which are barely developed, scrubber technology has done well to evolve as far as it has.

The following "challenges to industry" are given for the purpose of maximizing the reliability of any scrubber installation in the utility industry:

1) Utilities must assume responsibility to make the scrubber system work. If the decision is made to install scrubbers, the utility must accept the system as a developmental prototype and provide for an engineering and maintenance intensive installation. Provided with a proper design basis, the utility must have its own qualified staff of chemists as well as mechanical and chemical engineers to work with its architect/engineer and selected vendor in building the best system to meet site-specific conditions. Process guarantees and fixed-cost contracts are not sufficient to assure reliable, cost-effective scrubber operation. Plant personnel must give the scrubber operating and maintenance priority equal to all other power station systems.

2) Utilities should encourage vendors to achieve a mechanical "hybrid" system based on the extensive body of available operational experience, which incorporates the best components (i.e., absorber, mist eliminator, reheater, process equipment) of all presently available scrubber approaches. The goal of this effort is to standardize the scrubber hardware for low sulfur and high sulfur coals such that development priorities may be placed on chemical design.

From what we know today, some fundamental considerations for maximizing scrubber reliability include the following:

a) The SO_2 absorber should contain a minimum of internals to avoid plugging and scaling. Where particulates are to be removed in the scrubbing system, a two stage venturi/spray tower combination is recommended.

b) A vertical mist eliminator (horizontal gas flow) offers advantages of superior drainage, greater washing flexibility, and better adaptability to closed-loop demister wash water recycle. Horizontal mist eliminators appear to work satisfactorily for low sulfur coal and for high sulfur scrubbing with lime.

c) The most reliable, low maintenance reheater design appears to be blending heated ambient air with the flue gas. This approach is more capital intensive and less energy efficient, but is very reliable.

d) Scrubber booster fans appear to operate more reliably when they are dry. Forced draft fans and induced draft fans installed downstream from the reheater have encountered minimum corrosion problems.

e) Spare modules and backup components improve the ability of scrubbers to be cleaned and serviced while on-line.

3) Concentrate on scrubber process chemistry for site-specific design. Capability to predict chemical behavior and performance reliability before the scrubber is built is needed.

NON-REGENERABLE PROCESSES SESSION

Chairman: Michael A. Maxwell
Chief, Emissions/Effluent Technology Branch
Industrial Environmental Research Laboratory
U. S. Environmental Protection Agency
Research Triangle Park, North Carolina

Preceding page blank

IERL-RTP SCRUBBER STUDIES RELATED TO FORCED OXIDATION

Robert H. Borgwardt

Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina 27711

ABSTRACT

Small scale experiments indicate that forced oxidation of FGD scrubber sludge to gypsum is a feasible option for U.S. systems operating with high-sulfur Eastern coals containing chloride. Forced oxidation was carried out as an integral part of the first stage of a two stage system, by air sparging the effluent hold tank of the first stage at pH 4.5. Under these conditions complete limestone utilization was achieved, together with total oxidation of the calcium sulfite at atmospheric pressure. The performance of the oxidizer was limited by mass transfer of O_2 from the air bubbles to the liquor, consequently a tower containing a 5.5-m depth of slurry was used to maximize oxygen transfer efficiency. Complete oxidation of the sludge was thus obtained at an air stoichiometry of 2.6, without catalysts and without mechanical atomizers. The oxidized slurry settled at a rate ten times faster than that of sulfite slurries at normal oxidation levels. A settled sludge of increased density (60% solids) also resulted, a factor important to the environmental and economic aspects of sludge disposal. The combined effects of high utilization and oxidation can yield a 35% reduction in the amount of total wastes produced when the scrubber is used for fly ash collection. An important objective of these tests, to produce a sludge that is filterable to 80% solids, was not attained. This result, together with the observation that the settling properties can in some cases be dominated by the fly ash, favor the dry collection of fly ash ahead of the scrubber as an alternative method of reducing total waste.

Preceding page blank

IERL-RTP SCRUBBER STUDIES RELATED TO FORCED OXIDATION

INTRODUCTION

A typical 600-MW power plant, burning Eastern (3.5% S) coal, will generate 15 short tons of SO_2 per hour. When this amount of SO_2 is reacted with limestone to form an environmentally acceptable waste product, it yields as much as 100 tons of wet sludge per hour, exclusive of fly ash. The environmental and economic aspects of sludge disposal in quantities of this magnitude were cited at the previous FGD symposium as the major problem in application of throwaway processes. In view of the current preference on the part of the electric utility industry for non-regenerable FGD systems--in spite of the sludge problem--one may assume that the overall economics justify the expense and effort required to deal with it. EPA's assessment of the situation (Princiotta, 1975) pointed to the improvement of limestone utilization as one potential method of reducing the amount of sludge produced. It is also known that the single factor most influencing the amount of waste produced, per unit of SO_2 scrubbed, is the dewatering characteristics of the solid product. This property has been tentatively related to the degree of oxidation (Rossoff, 1974), by comparison of sludges produced in various full scale scrubbers.

Experiments at EPA's pilot plant at Research Triangle Park have been focused on this problem during the past year, both from the utilization and the oxidation standpoints. Increasing the utilization is, of course, an important objective for its own merits due to the effect on operating cost and the effect on mist eliminator performance (reduced fouling). High utilization is also considered, in our approach, to be a necessary condition for efficiently carrying out the oxidation step; i.e., the problems of achieving maximum utilization and oxidation are related. The ultimate objective was therefore to eliminate unreacted limestone as well as unoxidized calcium sulfite from the product solids. The purpose of this paper is to evaluate the prospects for achieving this objective with scrubbers operating at conditions that correspond to the use of high sulfur coal.

BACKGROUND

The feasibility of converting calcium sulfite scrubber slurry to gypsum was first indicated by I.C.I. in 1935; their original process design included an oxidation tower. That design, in which the oxidation step was carried out within the primary scrubber loop, was not efficient with respect to the amount of air required. Mitsubishi Heavy Industries reported (Uno, 1971) several improvements in the state of the art at the Second EPA Limestone Scrubbing Symposium. The MHI Lime/Gypsum Process employs two-stage scrubbing to adjust the pH to 4 - 4.5, followed by pressurized (3.8 - 4.8 atm) aeration using mechanically driven atomizers designed by the Japan Engineering & Consulting Co. (JECCO). The JECCO oxidizer, operating with catalysts and low pH, has been successfully applied in Japan to produce pure gypsum suitable for many industrial and construction uses. The suitability of this process to the operating conditions of U.S. scrubbers, which have both fly ash and chloride present in the slurry, has not been demonstrated. The experimental study at RTP was initiated primarily because of questions regarding these factors.

The MHI report showed that the mass transfer efficiency of the oxidizer, with respect to oxygen absorption, drops sharply at pH's greater than about 5.3 (with no Cl^- present). Fundamental kinetic studies conducted in the Soviet Union (Gladkii, 1974) showed a maximum rate for the oxidation of calcium sulfite slurries at pH 4.5. The HSO_3^- ion was identified as the active species undergoing oxidation and the homogeneous reaction rate was found to correlate directly with the concentration of that ion. Concurrent work carried out for EPA at Arthur D. Little, Inc. (Berkowitz, 1975) confirmed the Soviet results: oxidation rate is proportional to the bisulfite concentration in solution and, as a result, is an order of magnitude faster at pH 4 than at pH 7.

pH AND UTILIZATION

Achieving the low pH required for efficient oxidation of calcium sulfite slurry necessarily means that the scrubber must operate at a high level of limestone utilization. At equilibrium the slurry will have a pH of 6.4 as long as any unreacted limestone is present (Wen, 1975). The fact that pH's this high are rarely observed in practice is a strong clue that the system does not operate at equilibrium, even at normal limestone feed stroichiometry. As stoichiometry is reduced the concentration of unreacted limestone in the slurry is lowered and kinetic factors become totally dominant (recent tests at Shawnee with the scrubber operating at low stoichiometry showed that 90 hours of stirring was required to raise the slurry pH from 5.7 to 6.1). The loss of SO_2 removal efficiency under these conditions imposes a constraint on both the minimum pH and the maximum utilization that can be obtained with scrubbers of current design.

Techniques of optimizing the performance of limestone scrubbers at low stoichiometry--a necessary condition for improving utilization--have been examined at RTP and are reported elsewhere (Borgwardt, 1975) in detail. That study showed the kinetics of limestone dissolution/SO_2 precipitation can be accelerated by modifying the scrubber effluent hold tank to simulate tubular flow. Three or four stirred tanks, of 2 - 4 min. residence time each, were found suitable. The modified configuration (Figure 1) will achieve greater completion of the hold tank reactions in any given total residence time than a single stirred tank of the same residence time. It was shown that a single-loop scrubber operating at 18 - 20 cm water pressure drop could thus achieve about 90% utilization at 82% SO_2 removal efficiency. Ninety percent SO_2 removal efficiency and 94% limestone utilization were obtained by increasing the tower pressure drop (holdup) to 30 cm water. Under these conditions scrubber effluent pH's of 4.5 and lower were achieved. The relationship between pH, chloride, and the concentration of SO_2 in the scrubber liquor is shown in Figure 2 as observed in the RTP pilot plant.

Attempts to oxidize the purge slurry of a system operating in the scrubber configuration indicated by Figure 1 resulted in low conversion. Although the initial pH was less than 4.5, it increased during aeration to 6.4. It was evident that the unreacted limestone in the slurry was sufficient, even at utilizations of 90 - 94%, to slowly increase the pH so that efficient oxidation rates (and $CaSO_3 \cdot \frac{1}{2}H_2O$ dissolution rate) could not be maintained.

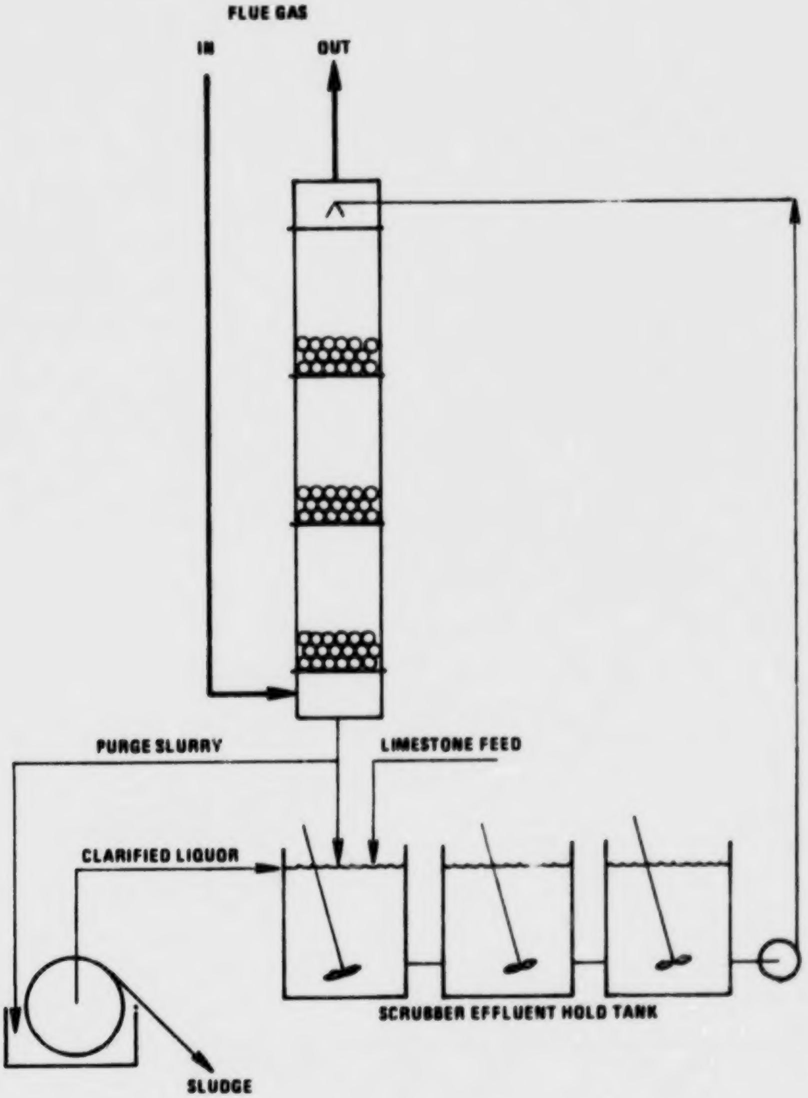

FLUE GAS
IN
OUT
PURGE SLURRY
LIMESTONE FEED
CLARIFIED LIQUOR
SCRUBBER EFFLUENT HOLD TANK
SLUDGE

Figure 1. Basic single-loop scrubbing configuration with E.H.T. modified to simulate tubular flow.

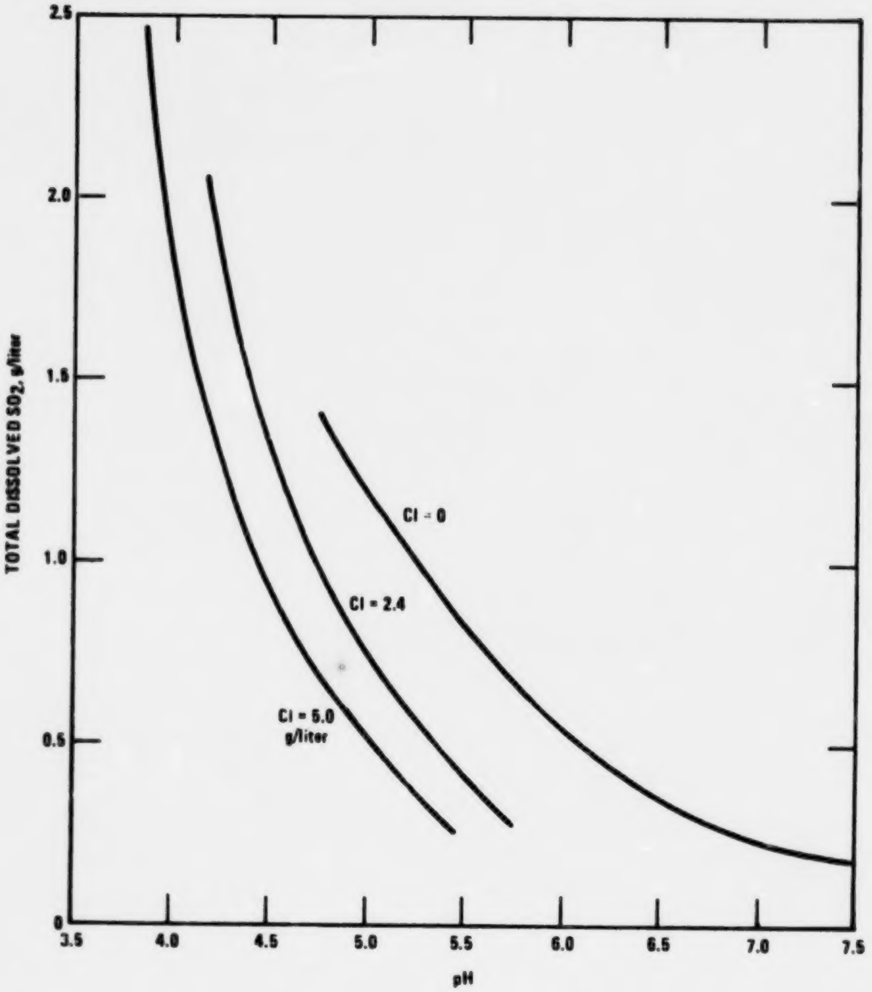


Figure 2. Concentration of dissolved sulfite in RTP scrubbing liquor as a function of pH.

The rise in pH is accounted for by the reaction:

$$2H^+ + CaCO_3 \rightarrow Ca^{++} + CO_2 + H_2O \qquad (1)$$

which consumes part of the H^+ needed to keep the reaction cycle, that is responsible for oxidation, going at constant pH:

$$HSO_3^- + \tfrac{1}{2}O_2 \rightarrow H^+ + SO_4^{--} \qquad (2)$$

$$H^+ + CaSO_3(s) \rightarrow Ca^{++} + HSO_3^- \qquad (3)$$

$$Ca^{++} + SO_4^{--} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O \qquad (4)$$

It was clear that additional H^+ must be provided during the course of reactions (2 - 3) if the pH was to be held constant.

STAGED SCRUBBING

The scrubber was modified to a two-stage configuration as shown in Figure 3 to permit a constant pH of 4.5 to be maintained in the oxidizer. A "staged" scrubber is defined here as one in which each gas-contacting stage has its own hold tank and slurry recirculation loop. Ideally, there would be no liquor carryover from the first to the second-stage loop by gas entrainment, and this was the case in all experiments reported here. The primary loop, in which most SO_2 absorption occurs, consisted of a TCA tower operating at 18 - 20 cm water ΔP, and was set up with multiple hold tanks. The purge stream from this loop, which would normally go to the clarifier or filter, is instead sent to the first stage loop where the slurry is contacted with additional SO_2--at maximum partial pressure--to provide the H^+ needed for reaction (1). The low pH required for oxidation and $CaSO_3(s)$ dissolution is thus maintained in the first stage with a minimum sacrifice of SO_2 removal efficiency in the primary scrubber. Since little additional SO_2 absorption is required in the first stage it was set up as a spray tower with low pressure drop (1 - 5 cm water) and low L/G. In full scale application the first stage would be a venturi; several systems, comprising a venturi in combination with a second, more efficient SO_2 scrubber, are already in operation in the U.S.

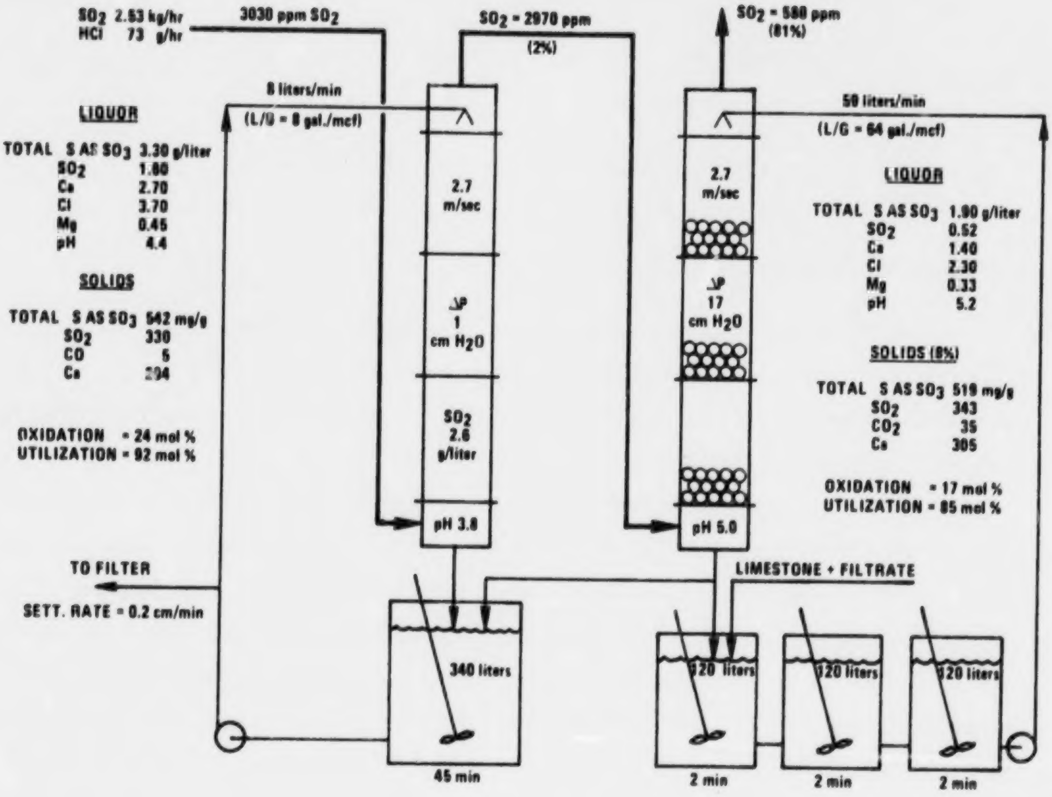

SO_2 2.53 kg/hr
HCl 73 g/hr
3030 ppm SO_2
SO_2 = 2970 ppm
(2%)
SO_2 = 580 ppm
(81%)
8 liters/min
(L/G = 8 gal./mcf)
58 liters/min
(L/G = 64 gal./mcf)
LIQUOR
TOTAL S AS SO_3 3.30 g/liter
SO_2 1.80
Ca 2.70
Cl 3.70
Mg 0.45
pH 4.4
SOLIDS
TOTAL S AS SO_3 542 mg/g
SO_2 330
CO 5
Ca 294
OXIDATION = 24 mol %
UTILIZATION = 92 mol %
2.7 m/sec
ΔP 1 cm H_2O
SO_2 2.6 g/liter
pH 3.8
2.7 m/sec
ΔP 17 cm H_2O
pH 5.0
LIQUOR
TOTAL S AS SO_3 1.90 g/liter
SO_2 0.52
Ca 1.40
Cl 2.30
Mg 0.33
pH 5.2
SOLIDS (8%)
TOTAL S AS SO_3 519 mg/g
SO_2 343
CO_2 35
Ca 305
OXIDATION = 17 mol %
UTILIZATION = 85 mol %
TO FILTER
SETT. RATE = 0.2 cm/min
LIMESTONE + FILTRATE
340 liters
45 min
120 liters
2 min
120 liters
2 min
120 liters
2 min

Figure 3. Staged scrubber without forced oxidation.

When forced oxidation is not attempted SO_2 builds up in the first stage scrubbing liquor to the concentrations shown in a typical run in this mode in Figure 3. In contrast, Figure 4 shows the result of forcing oxidation in the first stage by installing an air sparger in the hold tank. The accumulated SO_2 is oxidized in the liquor, together with the solid calcium sulfite that is fed from the primary loop. The latter is redissolved, oxidized in the liquid phase, and recrystallized as gypsum. Comparison of the liquor compositions shown in Figures 3 and 4 indicates that the steady-state SO_2 concentration in the first stage is no longer determined solely by pH and chloride when oxidation is forced: it is determined instead by the difference between the rate of oxidation and the rate of SO_2 feed to the oxidizer.

The pH was controlled at 4.5 in the oxidizer by adjusting the rate of limestone feed to the primary scrubbing loop. Since the utilization was already 85% in that loop, the additional SO_2 absorption in the first stage boosted the overall limestone utilization to values approaching 100%.

Figure 4 is the basic scrubber configuration used at RTP for forced oxidation. A major difference between this configuration and that used by MHI is the fact that oxidation was carried out as an integral part of the first stage in the RTP tests. This has the advantage of providing a slurry of essentially pure gypsum to control scaling in the first stage, which is most susceptible to that problem. We were thus able to operate at L/G = 1.3 l./m^3 (10 gal./mcf) in the first stage without scaling when set up according to Figure 4, but could not operate at this L/G when oxidation was not forced in the first stage. In the latter case gypsum scale formed in the upper parts of the tower where the feed slurry entered. Another advantage of our approach is the ability to control pH at the optimum value throughout the course of the oxidation reactions. This minimizes the stripping of SO_2 from the liquor in the oxidizer and increases the oxidation efficiency to the extent that it can be conducted at atmospheric pressure and without catalysts.

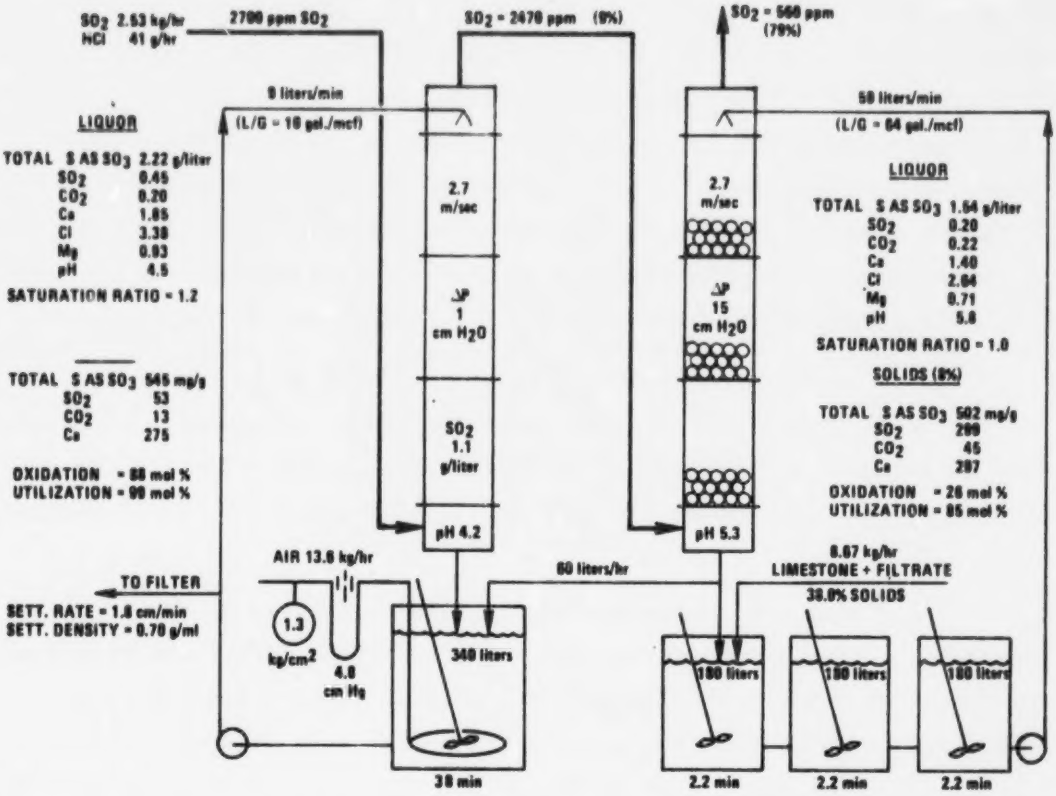

SO_2 2.53 kg/hr
HCl 41 g/hr
2700 ppm SO_2
SO_2 = 2470 ppm (9%)
SO_2 = 560 ppm (79%)
9 liters/min
(L/G = 16 gal./mcf)
LIQUOR
TOTAL S AS SO_3 2.22 g/liter
SO_2 0.45
CO_2 0.20
Ca 1.85
Cl 3.38
Mg 0.93
pH 4.5
SATURATION RATIO = 1.2
TOTAL S AS SO_3 545 mg/g
SO_2 53
CO_2 13
Ca 275
OXIDATION = 88 mol %
UTILIZATION = 98 mol %
2.7 m/sec
ΔP 1 cm H_2O
SO_2 1.1 g/liter
pH 4.2
2.7 m/sec
ΔP 15 cm H_2O
pH 5.3
58 liters/min
(L/G = 64 gal./mcf)
LIQUOR
TOTAL S AS SO_3 1.64 g/liter
SO_2 0.20
CO_2 0.22
Ca 1.40
Cl 2.84
Mg 0.71
pH 5.8
SATURATION RATIO = 1.0
SOLIDS (8%)
TOTAL S AS SO_3 502 mg/g
SO_2 209
CO_2 45
Ca 287
OXIDATION = 28 mol %
UTILIZATION = 85 mol %
TO FILTER
SETT. RATE = 1.8 cm/min
SETT. DENSITY = 0.70 g/ml
AIR 13.8 kg/hr
1.3 kg/cm^2
4.8 cm Hg
340 liters
38 min
60 liters/hr
8.67 kg/hr
LIMESTONE + FILTRATE
38.0% SOLIDS
180 liters
180 liters
180 liters
2.2 min
2.2 min
2.2 min

Figure 4. Staged scrubber with forced oxidation in the first stage (0.91-meter slurry depth; air stoichiometry = 6.3).

OXIDIZER EFFICIENCY

As indicated by the results shown in Figures 4 (no fly ash) and 5 (with fly ash), oxidation was successfully accomplished in the RTP scrubber by air sparging the first stage hold tank. The tank used for these initial tests was 71 cm in diameter and contained a 0.91-meter depth of slurry. Air was admitted at 0.7 - 1.4 kg/cm^2 pressure through a 50-cm diameter ring containing twenty-two 1.6 mm holes. Experience showed that the holes must point downward to prevent slurry from entering the tube, where the solids settle and eventually fill it. A 6.35-mm orifice was used to measure air flow. Water was added at 10 ml/min at a point ahead of the orifice to inhibit scale growth in the small holes of the sparge ring. Satisfactory results were obtained with sparge rings made of 316 stainless, PVC, and Teflon. The Teflon ring appeared best with respect to lack of corrosion and tendency to resist pluggage. The oxidizer tank was vented to the scrubber inlet so the air feed pressure represents sparge ring ΔP almost entirely.

Figures 4 and 5 show that the air feed rate required to complete the oxidation of the slurry with the type of oxidizer described above was 6 - 7 times the stoichiometric minimum, where air stoichiometry is defined by:

$$\text{air stoichiometry} = \frac{\text{g-atoms of oxygen fed to oxidizer}}{\text{g-moles of } SO_2 \text{ absorbed in scrubber}} = \frac{(\text{kg air fed/hr})\ 0.21\ (64.1)\ 2}{29\ (\text{kg } SO_2 \text{ fed/hr})\ \eta}$$

where η is the combined SO_2 absorption efficiency of both stages.

Attempts to reduce the air stoichiometry below 6 resulted in a buildup of HSO_3^- in the first stage scrubbing liquor, indicating that the rate of oxygen absorption had become the limiting factor--the rate of oxidation in the liquid phase at pH 4.5 has been shown (Gladkii, 1974) to increase with increasing HSO_3^- concentration; therefore, oxidation kinetics were not limiting. The HSO_3^- concentration rises in the first stage because, in this case, it is a dependent variable determined by the oxidation rate--and not vice versa. The steady-state

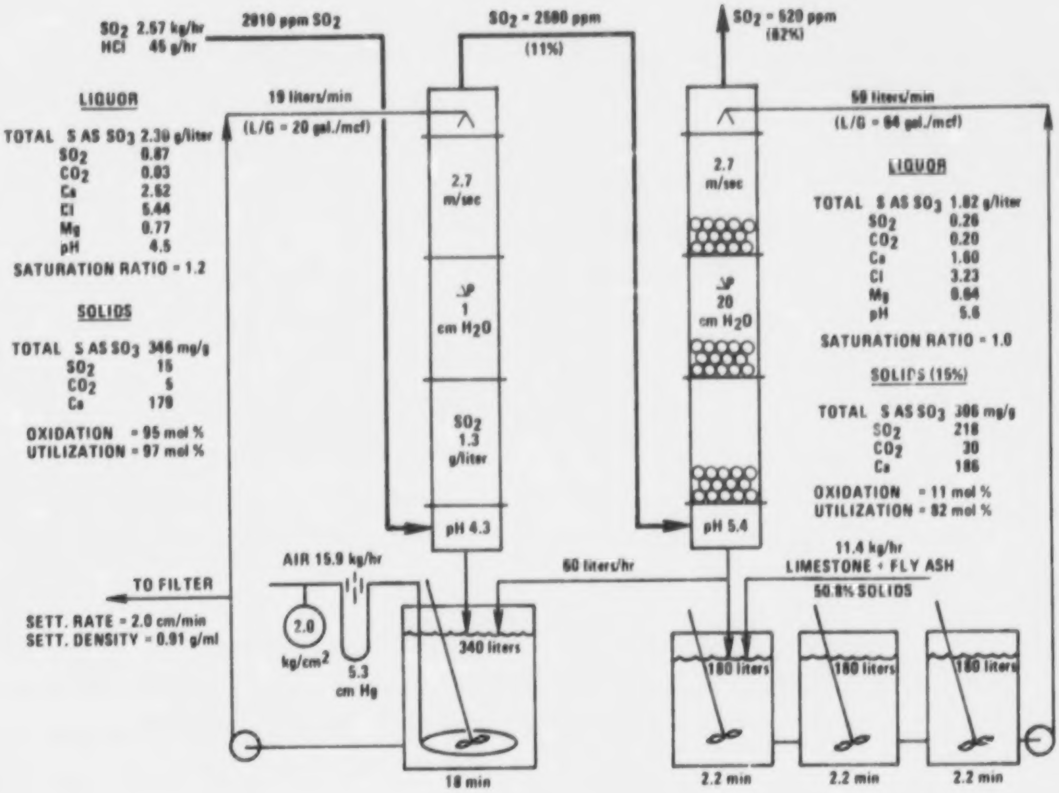

SO2 2.57 kg/hr
HCl 45 g/hr
2910 ppm SO2
SO2 = 2580 ppm
(11%)
SO2 = 520 ppm
(82%)
19 liters/min
(L/G = 20 gal./mcf)
59 liters/min
(L/G = 64 gal./mcf)
LIQUOR
TOTAL S AS SO3 2.30 g/liter
SO2 0.87
CO2 0.03
Ca 2.62
Cl 5.44
Mg 0.77
pH 4.5
SATURATION RATIO = 1.2
SOLIDS
TOTAL S AS SO3 346 mg/g
SO2 15
CO2 5
Ca 179
OXIDATION = 95 mol %
UTILIZATION = 97 mol %
2.7 m/sec
ΔP 1 cm H2O
SO2 1.3 g/liter
pH 4.3
2.7 m/sec
ΔP 20 cm H2O
pH 5.4
LIQUOR
TOTAL S AS SO3 1.82 g/liter
SO2 0.26
CO2 0.20
Ca 1.60
Cl 3.23
Mg 0.64
pH 5.6
SATURATION RATIO = 1.0
SOLIDS (15%)
TOTAL S AS SO3 306 mg/g
SO2 218
CO2 30
Ca 186
OXIDATION = 11 mol %
UTILIZATION = 82 mol %
AIR 15.9 kg/hr
2.0
kg/cm2
5.3 cm Hg
60 liters/hr
11.4 kg/hr
LIMESTONE + FLY ASH
50.8% SOLIDS
TO FILTER
SETT. RATE = 2.0 cm/min
SETT. DENSITY = 0.91 g/ml
340 liters
18 min
180 liters
180 liters
180 liters
2.2 min
2.2 min
2.2 min

Figure 5. Forced oxidation with fly ash addition
(0.91-m slurry depth; air stoichiometry = 7.0).

oxidation rate is limited, in turn, by the absorption of oxygen from the air bubbles. The number of bubbles per unit volume of slurry decreases in the oxidizer with the air feed rate and thus the air/liquor mass-transfer surface limits the amount of oxygen that can be absorbed during the time of contact between the bubbles and slurry.

The efficiency of oxygen mass transfer to a sulfite solution is discussed by Urza and Jackson (1975) and their rationale was followed to improve the oxidation efficiency of the RTP system. The absorption of O_2 from air bubbles of a given size was shown to be proportional to liquor depth (bubble residence time) for oxidizer heights to 16 meters. The oxidation kinetics in solution influence the overall rate only via the effect on the concentration of oxygen in the bulk liquid phase. From this viewpoint the high HSO_3^- concentration at low pH enhances the absorption of oxygen by minimizing the concentration of dissolved O_2 in the liquor.

Accordingly, the stirred-tank oxidizer was replaced with a tower (PVC pipe) having a static slurry depth of 3.2, and later, 5.5 meters. As shown in Figures 6, 7 and 8 these changes permitted us to reduce the air stoichiometry ultimately to 2.6. The air sparger used in the tower had the same size and number of holes as that used in the stirred tank. As a result of the increased bubble residence time afforded by greater slurry depth in the oxidizer, oxygen transfer efficiency was improved and more SO_2 was oxidized per kg of air injected. It is reasonable to expect that stoichiometries lower than 2.6 can be achieved by further increase of the oxidizer height. Also, as pointed out by Urza and Jackson, one can expect a bonus in the form of a reduced energy requirement for air compression, per unit of O_2 transferred.[8]

As indicated by Figure 6 the residence time of the slurry in the oxidizer (with respect to the rate of recirculation to the first stage tower) was reduced to 7 min. without ill effect on either the oxidation efficiency or scaling. The $CaSO_4 \cdot 2H_2O$ supersaturation of the liquor increased, however, from 1.1x at 16 min. to 1.3 - 1.4x at 7 min. residence time. The liquor saturation ratios shown in Figures 3 - 8, 11 were determined at 25° C by the direct test method.

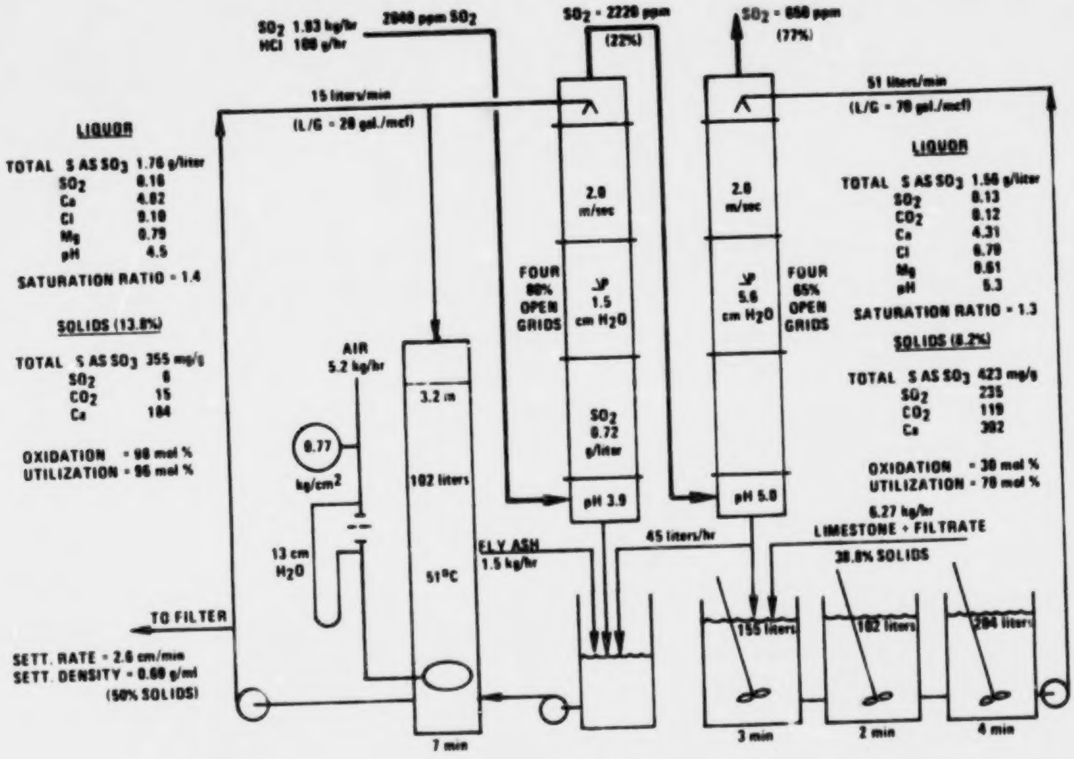

SO_2 1.93 kg/hr
HCl 100 g/hr
2040 ppm SO_2
SO_2 = 2220 ppm
(22%)
SO_2 = 650 ppm
(77%)
15 liters/min
(L/G = 20 gal./mcf)
51 liters/min
(L/G = 70 gal./mcf)
LIQUOR
TOTAL S AS SO_3 1.76 g/liter
SO_2 0.16
Ca 4.02
Cl 9.10
Mg 0.79
pH 4.5
SATURATION RATIO = 1.4
SOLIDS (13.8%)
TOTAL S AS SO_3 355 mg/g
SO_2 6
CO_2 15
Ca 184
OXIDATION = 98 mol %
UTILIZATION = 96 mol %
2.0 m/sec
FOUR 80% OPEN GRIDS
ΔP 1.5 cm H_2O
SO_2 0.72 g/liter
pH 3.9
2.0 m/sec
ΔP 5.6 cm H_2O
FOUR 65% OPEN GRIDS
pH 5.0
LIQUOR
TOTAL S AS SO_3 1.56 g/liter
SO_2 0.13
CO_2 0.12
Ca 4.31
Cl 6.70
Mg 0.61
pH 5.3
SATURATION RATIO = 1.3
SOLIDS (8.2%)
TOTAL S AS SO_3 423 mg/g
SO_2 235
CO_2 119
Ca 302
OXIDATION = 30 mol %
UTILIZATION = 70 mol %
AIR 5.2 kg/hr
0.77 kg/cm^2
3.2 m
102 liters
51°C
13 cm H_2O
FLY ASH 1.5 kg/hr
45 liters/hr
6.27 kg/hr
LIMESTONE + FILTRATE
38.8% SOLIDS
TO FILTER
SETT. RATE = 2.6 cm/min
SETT. DENSITY = 0.60 g/ml
(50% SOLIDS)
155 liters
102 liters
204 liters
7 min
3 min
2 min
4 min

Figure 6. Oxidizer depth increased to 3.2 meters (air stoichiometry = 3.2).

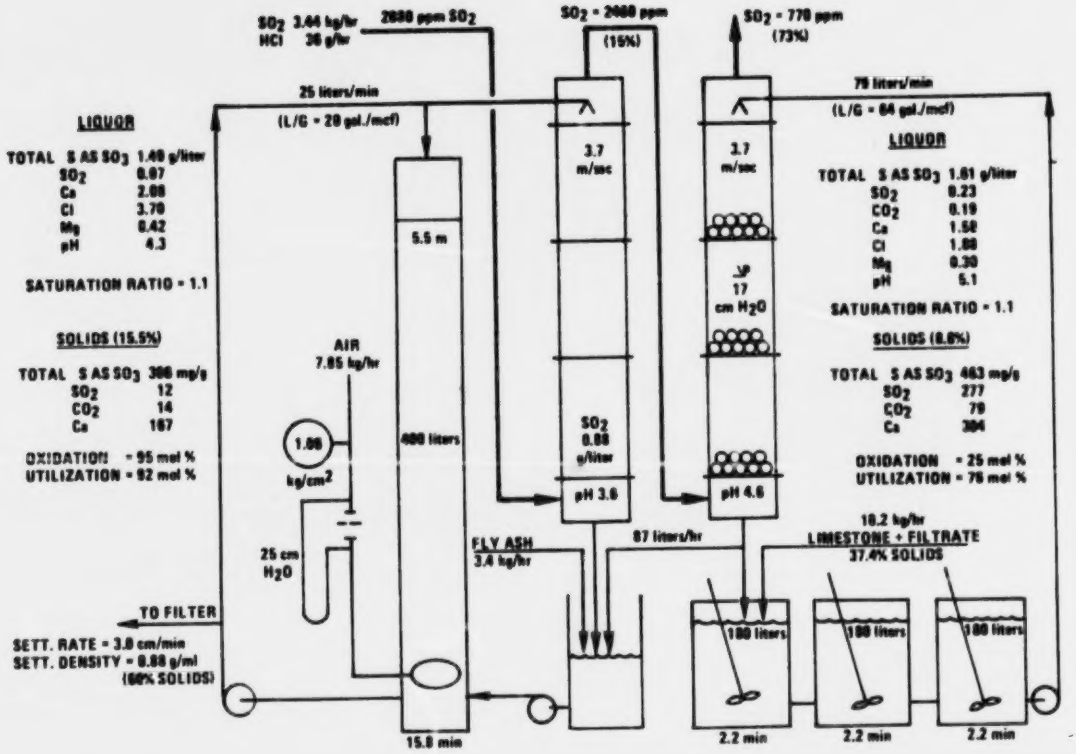

SO$_2$ 3.44 kg/hr
HCl 36 g/hr
2880 ppm SO$_2$
SO$_2$ = 2080 ppm
(15%)
SO$_2$ = 770 ppm
(73%)
25 liters/min
(L/G = 20 gal./mcf)
79 liters/min
(L/G = 64 gal./mcf)
LIQUOR
TOTAL S AS SO$_3$ 1.48 g/liter
SO$_2$ 0.07
Ca 2.08
Cl 3.70
Mg 0.42
pH 4.3
SATURATION RATIO = 1.1
SOLIDS (15.5%)
TOTAL S AS SO$_3$ 306 mg/g
SO$_2$ 12
CO$_2$ 14
Ca 167
OXIDATION = 95 mol %
UTILIZATION = 92 mol %
5.5 m
AIR
7.85 kg/hr
1.06
kg/cm^2
400 liters
25 cm
H$_2$O
3.7
m/sec
SO$_2$
0.88
g/liter
pH 3.6
3.7
m/sec
ΔP
17
cm H$_2$O
pH 4.6
LIQUOR
TOTAL S AS SO$_3$ 1.61 g/liter
SO$_2$ 0.23
CO$_2$ 0.19
Ca 1.58
Cl 1.88
Mg 0.30
pH 5.1
SATURATION RATIO = 1.1
SOLIDS (8.8%)
TOTAL S AS SO$_3$ 463 mg/g
SO$_2$ 277
CO$_2$ 79
Ca 304
OXIDATION = 25 mol %
UTILIZATION = 76 mol %
FLY ASH
3.4 kg/hr
87 liters/hr
10.2 kg/hr
LIMESTONE + FILTRATE
37.4% SOLIDS
TO FILTER
SETT. RATE = 3.0 cm/min
SETT. DENSITY = 0.88 g/ml
(60% SOLIDS)
15.8 min
180 liters
180 liters
180 liters
2.2 min
2.2 min
2.2 min

Figure 7. Oxidizer depth increased to 5.5 meters
(air stoichiometry = 2.9).

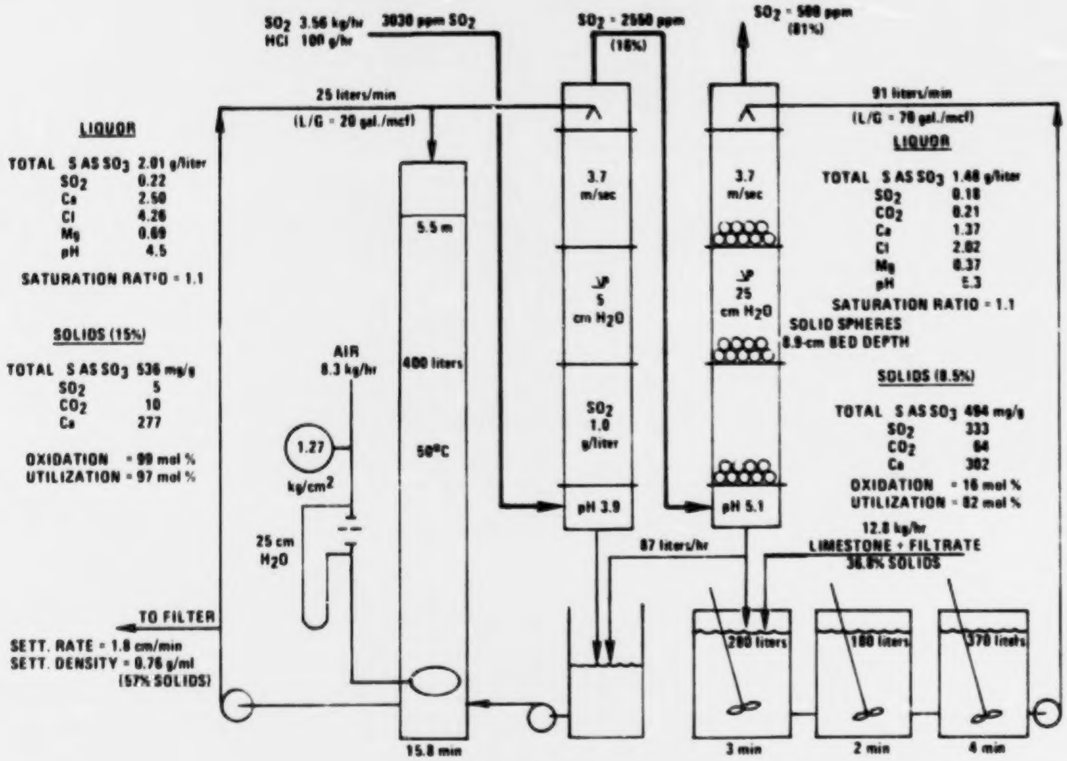

SO2 3.56 kg/hr
HCl 100 g/hr
3030 ppm SO2
SO2 = 2560 ppm
(16%)
SO2 = 580 ppm
(81%)
25 liters/min
(L/G = 20 gal./mcf)
91 liters/min
(L/G = 70 gal./mcf)
LIQUOR
TOTAL S AS SO3 2.01 g/liter
SO2 0.22
Ca 2.50
Cl 4.26
Mg 0.69
pH 4.5
SATURATION RAT'O = 1.1
SOLIDS (15%)
TOTAL S AS SO3 536 mg/g
SO2 5
CO2 10
Ca 277
OXIDATION = 99 mol %
UTILIZATION = 97 mol %
5.5 m
400 liters
50°C
AIR
8.3 kg/hr
1.27
kg/cm2
25 cm
H2O
3.7
m/sec
ΔP
5
cm H2O
SO2
1.0
g/liter
pH 3.9
3.7
m/sec
ΔP
25
cm H2O
pH 5.1
LIQUOR
TOTAL S AS SO3 1.48 g/liter
SO2 0.18
CO2 0.21
Ca 1.37
Cl 2.02
Mg 0.37
pH 5.3
SATURATION RATIO = 1.1
SOLID SPHERES
8.9-cm BED DEPTH
SOLIDS (8.5%)
TOTAL S AS SO3 494 mg/g
SO2 333
CO2 64
Ca 302
OXIDATION = 16 mol %
UTILIZATION = 82 mol %
12.8 kg/hr
LIMESTONE + FILTRATE
36.8% SOLIDS
87 liters/hr
TO FILTER
SETT. RATE = 1.8 cm/min
SETT. DENSITY = 0.76 g/ml
(57% SOLIDS)
280 liters
180 liters
370 liters
15.8 min
3 min
2 min
4 min

Figure 8. No fly ash addition (5.5-m oxidizer depth; air stoichiometry = 2.6).

OXIDATION AND SETTLING RATE

An important objective of forced oxidation is to improve the settling rate of the slurry and reduce the size of the clarifier which, at Shawnee for example, is the largest single vessel in the scrubber system. An improvement in the settling rate as a result of oxidation was first reported by TVA (Kelso, 1971), an effect attributed to enlargement of the crystal size. Data accumulated at RTP during the course of this study (Figure 9) confirm TVA's conclusion. Using Shawnee (Fredonia) limestone, the fully oxidized slurry settled 10 times faster than sulfite slurries in the normal range of oxidation. Our results also confirm that very high levels of oxidation must be attained before the improvement becomes apparent.

SETTLED DENSITY AND SLUDGE PRODUCTION

The Aerospace sludge disposal study (Rossoff, 1974) has pointed out the importance of sludge density to its environmental acceptability. This property is the major determinant of compaction strength, compressibility, and permeability. One of the reasons for the interest in forced oxidation is the possibility of producing a sludge that can be dewatered by filtration to a moisture content near the optimum value for compaction. Thus, the use of dry additives, that are required to attain the optimum water content with sulfite sludge, might be avoided. The load-bearing strength and drainability are both superior for a compacted sulfate sludge, and a waste product in this form is more suitable than sulfite sludge for direct disposal as landfill.[2]

Figure 10 shows the density of settled sludges obtained at RTP, which clearly increased with oxidation to a value of 0.9 g dry solids per ml of sludge volume (equivalent to 60% solids) when fully oxidized. In actual practice, settled densities corresponding to 35 - 45% solids are obtained with most sulfite sludges.[2] The values of Figure 10 are somewhat lower because the test was made under conditions that avoid compression of the settled solids, and they were not mechanically disturbed by a clarifier rake. It is not unlikely therefore that a

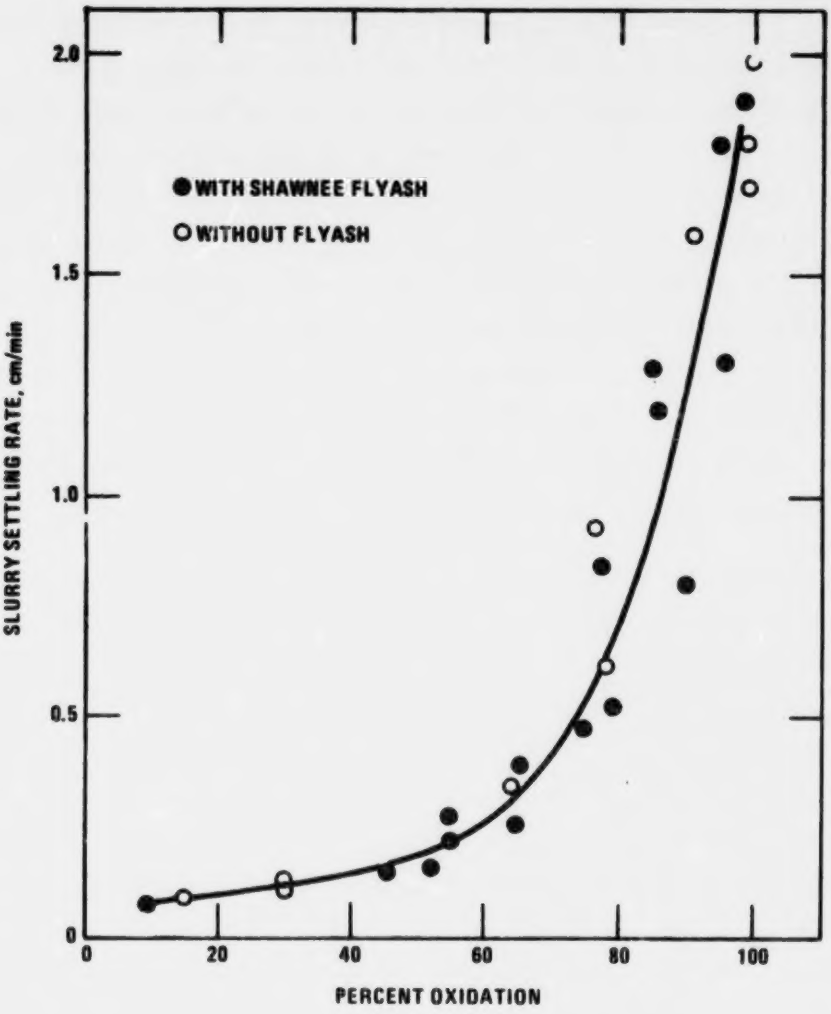


Figure 9. Settling rate of RTP limestone scrubber slurries as a function of oxidation (50° C).

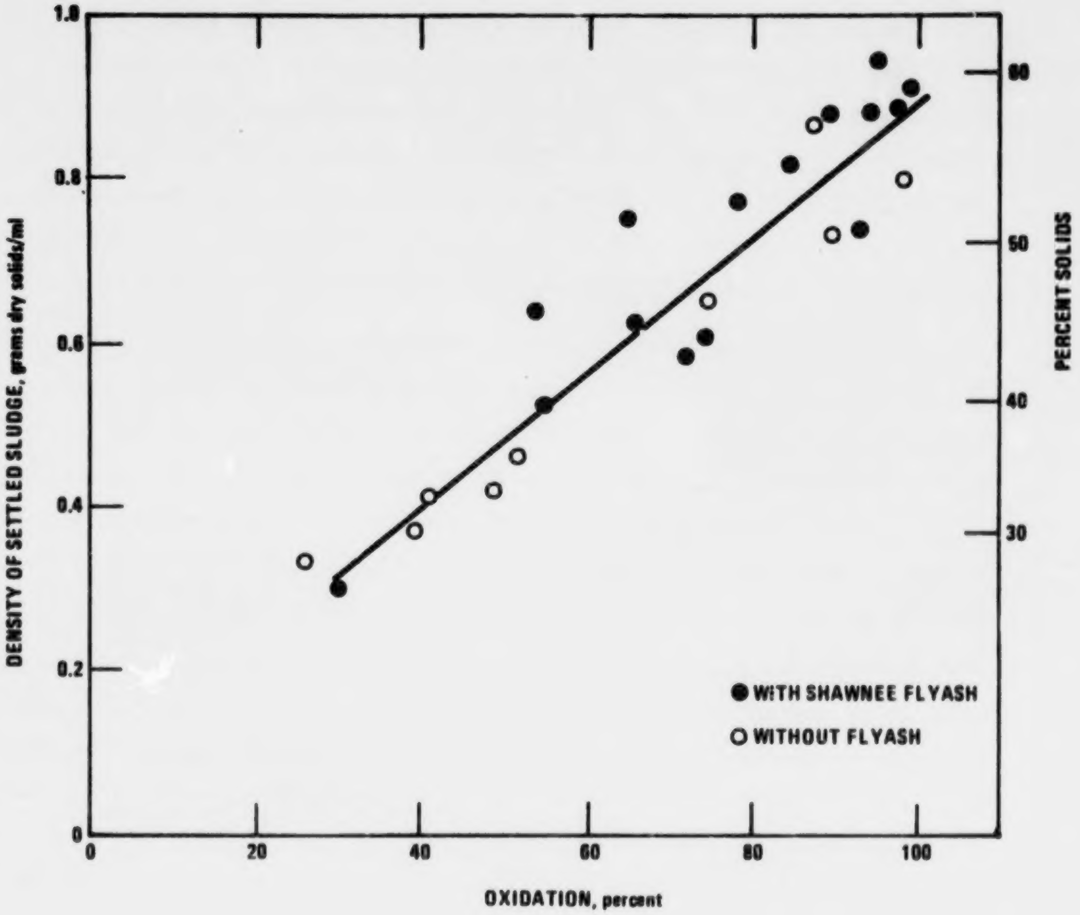


Figure 10. Settled (quiescent) density of RTP limestone scrubber sludge, as a function of oxidation.

large clarifier would yield a fully oxidized sludge of 65% solids as expected.[2] Vacuum filtration of the oxidized slurry yielded a filter cake that averaged about 67% solids in our tests. These samples were reslurried in acetone and dried at room temperature to ensure that no water of hydration was lost. Thus, the objective of achieving 80% solids by filtration does not appear attainable on the basis of RTP results. This value is ideal for optimum compaction[2] and, therefore, is important to the question of direct disposal as a landfill.

It was also apparent from our tests that settled sludges containing 60% solids could only be obtained with the longer oxidizer residence time (16 min., based on the slurry recirculation rate to the first stage), which corresponds to about 5 hours residence time for the slurry in the first stage. At shorter residence time (7 min.) we obtained only 50% solids. This suggests that the additional crystal growth that can occur at long residence time is important to achieving maximum settled density.

The combined effects of high limestone utilization and improved dewatering properties of the sludge can be expected to have a significant impact upon the rate of waste production by a power plant. The effects are shown in Table 1 for various scrubber operating conditions, using the present situation defined in the SOTSEP report[1] as a basis for comparison. It is clear from Table 1 that: a) about 24% reduction of the amount of sludge can be obtained using standard gravity settling; b) a greater reduction of sludge is achieved by the use of more efficient dewatering equipment, such as vacuum filtration, than is possible by forced oxidation alone--thus, forced oxidation is justified only if gravity settling is used for dewatering.

FLY ASH

In general, settling rate is determined by the size of the smallest particles in suspension. It is clear from Figure 9 that the settling rate of the mixture of fly ash and $CaSO_4/CaSO_3$ was dominated by calcium sulfite. The fly ash fed to the scrubber during these tests, fly ash "A", had a settling rate of 5.5 cm/min at 50° C in the absence of $CaSO_4/CaSO_3$. Later tests with a different fly ash (fly ash "B", also

Table 1. ANNUAL WASTE SLUDGE PRODUCTION BY A 1000 MW COAL-FIRED POWER PLANT EQUIPPED WITH LIMESTONE FGD SCRUBBERS: FLY ASH COLLECTED IN SCRUBBER

(SHORT TONS)

Dewatering Procedure	Settling	Settling	Oxidation/Settling	Filtration	Oxidation/Filtration
Scrubber Operating Conditions	Utilization = 60% Oxidation = 10%	Utilization - 90% Oxidation = 10%	Utilization = 100% Oxidation = 100%	Utilization = 90% Oxidation = 10%	Utilization = 100% Oxidation = 100%
Coal Ash, Dry	338,000	338,000	338,000	338,000	338,000
$CaSO_3 \cdot \frac{1}{2}H_2O$	322,000	322,000	0	322,000	0
$CaSO_4 \cdot 2H_2O$	48,000	48,000	477,000	48,000	477,000
$CaCO_3$	185,000	30,800	0	30,800	0
Total Dry Solids	893,000	739,000	815,000	739,000	815,000
Solids Moisture, %	50	50	40	38	30[b]
Total Wet Sludge	1,790,000	1,478,000	1,482,000	1,192,000	1,164,000
Reduction of Sludge Production	Base Case[a]	17%	24%	33%	35%

[a] SOTSEP Report[1], p. 60

[b] Minimum moisture content obtainable by filtration in RTP tests

from the Shawnee Test Facility) showed that the settling characteristics of the scrubber slurry can be dominated by the fly ash instead of calcium sulfite. Fly ash "B" settled at 1.5 cm/min in water at 50° C and in the scrubber yielded a slurry that settled at a rate that was independent of oxidation. Not only was the settling rate of this slurry determined by the fly ash, but an adverse interaction was also apparent; e.g., a sample of fully oxidized slurry taken from the scrubber when operating without fly ash (and settling at a rate of 3 cm/min) settled at only 0.7 cm/min when it was reslurried with fly ash "B" and compared at the same temperature of 50° C. The settled density was also reduced by the presence of fly ash "B", the fully oxidized sludge containing only about 50% solids. In view of this experience it appears that the current industry trend toward dry collection of the fly ash prior to the scrubber is well advised. Although the pervasiveness of "B"-type fly ashes is not known, it is clear that minimal benefit can be expected from forced oxidation when it is present in the scrubber slurry.

Dry fly ash collection can be considered as an alternative to forced oxidation as a means of reducing the waste production and, if it is mixed with the scrubber sludge, as a means of increasing the density. In view of the results at RTP, it may be the only way to achieve the desired 80% solids for optimum compaction. Table 2 summarizes the relative amounts of waste produced under varying conditions when the fly ash is collected ahead of the scrubber. Comparison of Tables 1 and 2 shows that the reduction of total waste is greater for dry fly ash collection than is possible when the ash is collected in the scrubber, and this is true whether or not forced oxidation is employed. As indicated by Table 2, however, blending the dry fly ash with the scrubber sludge can also yield the desired 80% solids if the sludge is oxidized.

SO_2 ABSORPTION

As indicated in Figure 6 the scrubber was successfully operated with both stages set up as spray towers at a total pressure drop of only 8 cm water and 77 - 80% SO_2 removal. When an L/G of 2.7 $l./m^3$

Table 2. ANNUAL WASTE SLUDGE PRODUCTION BY A 1000 MW COAL-FIRED POWER PLANT EQUIPPED WITH LIMESTONE FGD SCRUBBERS: DRY FLY ASH COLLECTION

(SHORT TONS)

Dewatering Procedure	Settling	Settling	Oxidation/Settling	Oxidation/Filtration	Filtration
Scrubber Operating Conditions	Utilization = 60% Oxidation = 10%	Utilization = 90% Oxidation = 10%	Utilization = 100% Oxidation = 100%	Utilization = 100% Oxidation = 100%	Utilization = 90% Oxidation = 10%
Coal Ash, Dry	338,000	338,000	338,000	338,000	338,000
$CaSO_3 \cdot \frac{1}{2}H_2O$	322,000	322,000	0	0	322,000
$CaSO_4 \cdot 2H_2O$	48,000	48,000	477,000	477,000	48,000
$CaCO_3$	185,000	30,800	0	0	30,800
Solids Moisture, %	50	50	40	30	38
Total Wet Sludge	1,110,000	817,000	795,000	682,000	647,000
Total Waste	1,448,000	1,155,000	1,133,000	1,020,000	985,000
Reduction of Waste Compared to Base Case	19%	35%	37%	43%	45%
Final Sludge [a] Density, % Solids	63	66	73	80	75

[a] Assuming dry fly ash is blended with wet sludge.

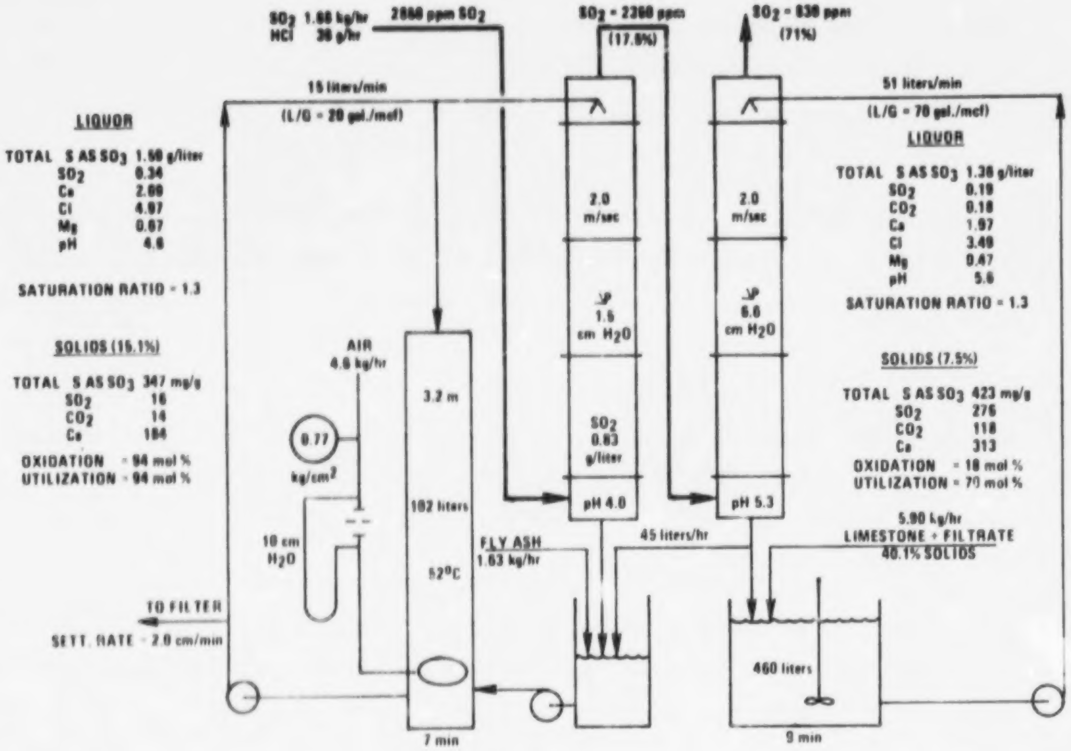
(20 gal./mcf) was maintained in the first stage, the SO_2 removal in that stage averaged 20% at a gas velocity of 2.0 m/sec. Thus, a utilization of only 70-80% is required in the second stage in order to control pH at the desired value, and this range of utilization is within the capability of a single stirred tank. Figure 11 summarizes a test in which one 9-min. hold tank was used in the second stage. Although a reduction of the overall SO_2 removal efficiency (to 71%) occurred, the test showed that complete oxidation could still be obtained.

Solid spheres were used in most of the tests at 3.7 m/sec gas velocity to permit the L/G to be increased above the 6.7 $1./m^3$ (50 gal./mcf) that is normally the upper limit for hollow (5-g) spheres. The solid (linear polyethylene) spheres were 2.5 cm diameter and formed three beds, each bed 7.6 cm deep. This tower could be operated at L/G's as high as 12 $1./m^3$ without flooding at 3.7 m/sec. The improvement results from the reduction of sphere buoyancy, which causes the light spheres to congregate against the upper retaining grid. An alternate solution is to increase the grid spacing, which in our case was 91 cm. As shown in Figures 7 and 8 the solid spheres gave 73% removal at L/G = 8.5 $1./m^3$, 17 cm water ΔP and 81% removal at L/G = 9.4 $1./m^3$, 25 cm ΔP, when operating at 3.7 m/sec.

CONCLUSIONS AND RECOMMENDATIONS

The high oxidation efficiency obtained by air sparging at atmospheric pressure indicates that considerable improvement in performance of forced oxidation systems can be realized if the oxidation step is carried out as an integral part of the first stage. The improvement will permit the use of simpler, less costly oxidizer designs, will reduce the power requirements and will eliminate the need for catalysts. The air stoichiometry is determined primarily by the height of the oxidizer and future tests should be made with slurry depths to at least 18 meters. Stoichiometries below 2.6 can thus be expected, and a minimum power for air compression.

It is clear that many variations of the two-stage configuration are workable, including two spray towers, multiple or single hold tanks, and



SO_2 1.66 kg/hr
HCl 36 g/hr
2880 ppm SO_2
SO_2 = 2360 ppm
(17.8%)
SO_2 = 830 ppm
(71%)
15 liters/min
(L/G = 20 gal./mcf)
51 liters/min
(L/G = 70 gal./mcf)
LIQUOR
TOTAL S AS SO_3 1.50 g/liter
SO_2 0.34
Ca 2.09
Cl 4.07
Mg 0.07
pH 4.6
SATURATION RATIO = 1.3
SOLIDS (15.1%)
TOTAL S AS SO_3 347 mg/g
SO_2 16
CO_2 14
Ca 184
OXIDATION = 94 mol %
UTILIZATION = 94 mol %
2.0 m/sec
ΔP 1.5 cm H_2O
SO_2 0.83 g/liter
pH 4.0
2.0 m/sec
ΔP 6.6 cm H_2O
pH 5.3
LIQUOR
TOTAL S AS SO_3 1.38 g/liter
SO_2 0.19
CO_2 0.18
Ca 1.97
Cl 3.49
Mg 0.47
pH 5.6
SATURATION RATIO = 1.3
SOLIDS (7.5%)
TOTAL S AS SO_3 423 mg/g
SO_2 276
CO_2 118
Ca 313
OXIDATION = 18 mol %
UTILIZATION = 70 mol %
AIR 4.6 kg/hr
0.77
kg/cm^2
10 cm H_2O
3.2 m
102 liters
52°C
7 min
FLY ASH 1.63 kg/hr
45 liters/hr
5.90 kg/hr
LIMESTONE + FILTRATE
40.1% SOLIDS
460 liters
9 min
TO FILTER
SETT. RATE = 2.0 cm/min

Figure 11. Single stirred tank in primary scrubber loop.

gas velocities ranging from 2.0 - 3.7 m/sec (6.7 - 12 fps). Our results favor the following conditions: 3.7 m/sec gas velocity, minimum ΔP in the first stage at L/G = 1.3 - 2.7 1./m^3, maximum pressure drop (hold up) in the second stage, and multiple hold tanks of 6 - 9 min total residence time in the second stage. The oxidizer should have a residence time of 15 min to provide a settled sludge of maximum density.

The benefits to be derived from oxidation, in terms of improvement of the physical properties of the waste, will depend upon the properties of the fly ash as well as the properties of the gypsum produced. In any case our results do not indicate that a sludge filterable to more than 70% solids can be expected. Dry fly ash collection would, under these circumstances, be the preferred method of minimizing waste production as it would afford the only opportunity to produce a directly disposable landfill.

The chemical properties of gypsum, vis-a-vis calcium sulfite as the most desired waste product, is the principal issue upon which the final decision must be made as to whether forced oxidation is to be undertaken on a large scale in the U.S.

REFERENCES

1. Princiotta, F.T., "Sulfur Oxide Throwaway Sludge Evaluation Panel (SOTSEP): Final Report, Volume II" EPA-650/2-75-010-b (NTIS No. PB 242-619/AS), April 1975.

2. Rossoff, J., et al., "Disposal of By-Products from Non-Regenerable Flue Gas Desulfurization Systems," Proceedings: Symposium on Flue Gas Desulfurization-Atlanta Nov. 1974, Volume I" EPA-650/2-74-126-a (NTIS No. PB 242-572/AS), pp. 399-441, December 1974.

3. Uno, T., et al., "The Pilot Scale R&D and Prototype Plant of MHI Lime-Gypsum Process," Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, pp. 834-849, November 1971.

4. Gladkii, A.V., et al., "State Scientific Research Institute of Industrial and Sanitary Gas Cleaning (Moscow)," Report for Protocol Point A-1, Development of Lime/Limestone Scrubbing for Stack Gas Desulfurization, US/USSR Sulfur Oxides Technology Sub-Group, 1974.

5. Berkowitz, J.B., et al., "Scale Control in Limestone Wet Scrubbing Systems," EPA-650/2-75-031 (NTIS No. PB 243-309/AS), pp. 52-56, April 1975.

6. Wen, C.Y., et al., Ibid., p. 66, April 1975.

7. Borgwardt, R.H., "Increasing Limestone Utilization in FGD Scrubbers," paper presented at 68th Annual Meeting, AIChE. Los Angeles Nov. 1975.

8. Urza, I.J., and Jackson, M.L., "Pressure Aeration in a 55-ft Bubble Column," Ind. Eng. Chem., Process Des. Dev., 15 pp. 106-113, April 1975.

9. Kelso, T.M., et al., "Limestone Wet-Scrubbing Pilot Plant at Colbert Steam Plant," Process Engineering Branch Report, Tennessee Valley Authority, April 1971.

RESULTS OF MIST ELIMINATION AND ALKALI UTILIZATION TESTING AT THE EPA ALKALI SCRUBBING TEST FACILITY

M. Epstein, H. N. Head, S. C. Wang, and D. A. Burbank

Bechtel Corporation
50 Beale Street
San Francisco, California 94105

ABSTRACT

Testing with a single stage, 3-pass, open-vane, 316 stainless steel chevron mist eliminator in both the venturi/spray tower and TCA systems with lime and limestone has shown that the accumulation of soft solids on the mist eliminator is a strong function of alkali utilization. For high alkali utilization (greater than about 85 percent), the mist eliminator was kept free of solids deposits by use of intermittent fresh water wash on both the top and bottomside. For alkali utilization less than about 85 percent, intermittent top and bottomside wash with fresh water did not limit solids accumulation. However, for these conditions, a continuous bottom wash with diluted clarified liquor used in combination with an intermittent topside fresh water wash was shown to limit soft solids buildup to less than 10 percent restriction within the mist eliminator.

In the venturi/spray tower system with lime additive, reliable long-term operation of the chevron mist eliminator with intermittent top and bottomside fresh water wash was demonstrated during constant load operation at the maximum attainable spray tower gas velocity of 9.4 ft/sec and during variable load (cycling gas rate) operation. The lime utilization for these tests was about 90 percent. In the TCA system with limestone additive, reliable long-term operation of the chevron mist eliminator with intermittent top and bottomside fresh water wash was demonstrated at 12.5 ft/sec gas velocity at high alkali utilization (greater than about 90 percent). For both the lime and limestone testing, only about half the makeup water available during closed liquor loop operation was needed to wash the mist eliminators.

Utilization with lime in the venturi/spray tower system was normally greater than 85 percent for scrubber inlet liquor pH less than 9.0. Utilization with limestone in both scrubber inlet liquor pH of 6.0 to about 95 percent at a scrubber inlet pH of 5.2. Operation at reduced scrubber liquor pH, however, inherently causes a reduction in SO_2 removal efficiency. For the venturi/spray tower system with a single effluent hold tank, limestone utilization was not affected

Preceding page blank

by a change in residence time from 20 to 12 minutes. Limestone utilization declined, however, at 6 minutes residence time for scrubber inlet liquor pH greater than about 5.6. For the TCA system at 12 minutes total residence time and at scrubber inlet liquor pH greater than about 5.0, higher limestone utilization was achieved with three hold tanks in series than with a single hold tank.

CONTENTS

Section Page

1 INTRODUCTION 149

2 ADVANCED TEST PROGRAM OBJECTIVES AND SCHEDULE 157

3 LIME TESTING WITH A CHEVRON MIST ELIMINATOR IN THE VENTURI/SPRAY TOWER SYSTEM 161

- 3.1 Testing at Constant Gas Velocity 161
- 3.2 Variable Load Testing 168
- 3.3 Conclusions 170

4 LIMESTONE TESTING WITH A WASH TRAY AND CHEVRON MIST ELIMINATOR IN SERIES IN THE TCA SYSTEM 172

- 4.1 Testing at 8.6 ft/sec Superficial Gas Velocity 172
- 4.2 Testing at 10 and 12 ft/sec Superficial Gas Velocity 175
- 4.3 Conclusions 176

5 LIMESTONE UTILIZATION TESTING IN THE VENTURI/SPRAY TOWER AND TCA SYSTEMS 177

- 5.1 Utilization Testing in the Venturi/Spray Tower System with Variable Residence Time 178
- 5.2 Utilization Testing in the TCA System with Three Hold Tanks in Series 184
- 5.3 Conclusions 193

6 EFFECT OF ALKALI UTILIZATION ON MIST ELIMINATOR OPERABILITY 196

- 6.1 Summary of Mist Eliminator Operability During Lime and Limestone Reliability Testing 196

Section **Page**

6.2 Mist Eliminator Operability During Limestone Utilization Testing 198
6.3 Conclusions 203

7 REFERENCES 204

Section 1

INTRODUCTION

In June 1968, the EPA[*] initiated a program to test prototype lime and limestone wet-scrubbing systems for removing sulfur dioxide and particulates from coal-fired boiler flue gases. The program was conducted in a test facility integrated into the flue gas ductwork of boiler No. 10 at the Tennessee Valley Authority (TVA) Shawnee Power Station, Paducah, Kentucky. Bechtel Corporation of San Francisco was the major contractor and test director, and TVA was the constructor and facility operator.

The results of testing at the facility during the original program, which lasted from March 1972 to May 1974, are presented in References 1 and 2. The most significant reliability problem encountered during the testing was associated with scaling and/or plugging of mist elimination surfaces.

In June 1974, the EPA, through its Office of Research and Development and Industrial Environmental Research Laboratory[**] initiated a three-year advanced test program at the Shawnee facility. Bechtel Corporation is continuing as the major contractor and test director, and TVA as the constructor and facility operator. The major goals established for the advanced program are: (1) to continue long-term testing with emphasis on demonstrating reliable operation of the mist elimination systems at increased gas velocity, (2) to investigate advanced process and equipment design variations for improving system reliability and process economics, and (3) to perform long-term (2 to 5 month) reliability testing on promising process and equipment design variations.

---

[*] The National Air Pollution Control Administration prior to 1970.

[**] The Control Systems Laboratory prior to 1975.

Two parallel scrubbing systems are being operated during the advanced program:

- A venturi followed by a spray tower
- A Turbulent Contact Absorber (TCA)

Each system has its own slurry handling facilities and is capable of treating approximately 30,000 acfm of flue gas from the TVA Shawnee coal-fired boiler No. 10. This gas rate is equivalent to approximately 10 Mw of power plant generating capacity. Boiler No. 10 normally burns a high-sulfur bituminous coal which produces SO_2 concentrations of 1500 to 4500 ppm and inlet particulate loadings of 2 to 4 grains/scf in the flue gas.

Figures 1-1 and 1-2 (drawn with major dimensions to scale) show the two scrubber systems along with the mist elimination systems currently used for de-entraining slurry in the exit gas streams. The cross-sectional area of the spray tower is 50 ft^2 in both the scrubbing section and mist elimination section. The cross-sectional area of the TCA is 32 ft^2 in the scrubbing section and 49 ft^2 in the mist elimination section.

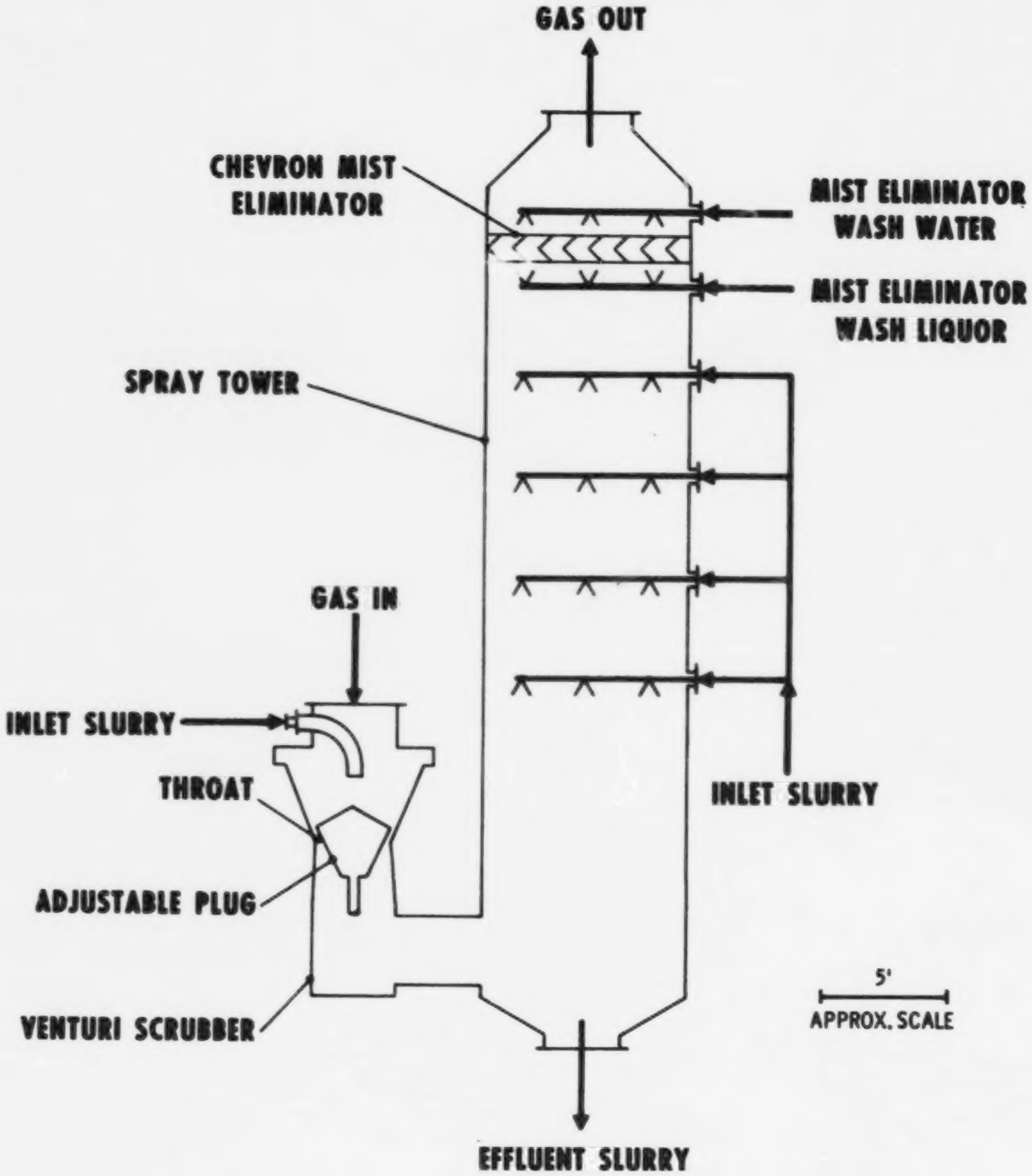

GAS OUT
CHEVRON MIST ELIMINATOR
MIST ELIMINATOR WASH WATER
MIST ELIMINATOR WASH LIQUOR
SPRAY TOWER
GAS IN
INLET SLURRY
THROAT
INLET SLURRY
ADJUSTABLE PLUG
5'
APPROX. SCALE
VENTURI SCRUBBER
EFFLUENT SLURRY

Figure 1-1. Schematic of Venturi Scrubber and Spray Tower

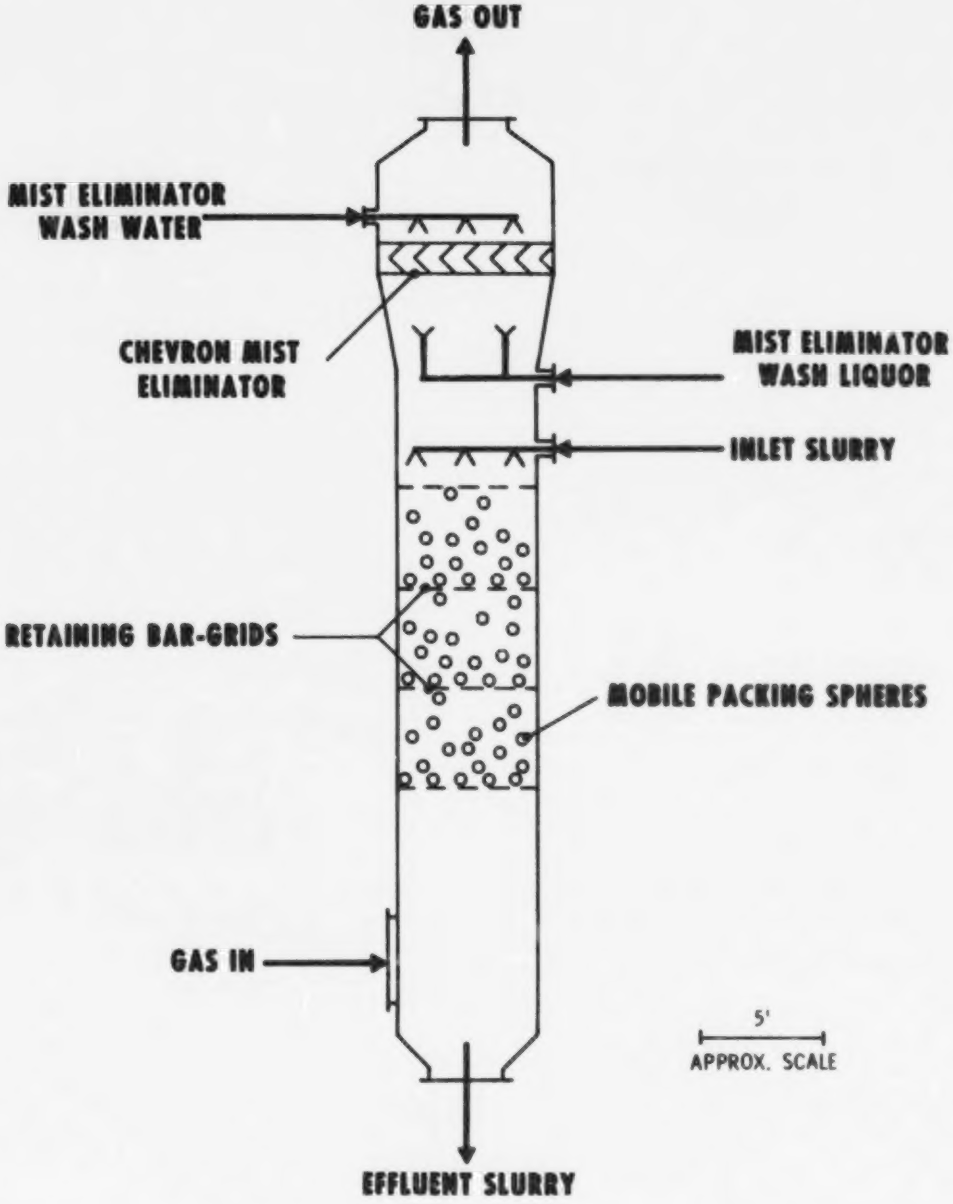

GAS OUT
MIST ELIMINATOR WASH WATER
CHEVRON MIST ELIMINATOR
MIST ELIMINATOR WASH LIQUOR
INLET SLURRY
RETAINING BAR-GRIDS
MOBILE PACKING SPHERES
GAS IN
5'
APPROX. SCALE
EFFLUENT SLURRY

Figure 1-2. Schematic of Three-Bed TCA Scrubber

The TCA utilizes a fluidized bed of 1 1/2 inch diameter, light weight (5.0-6.5 gram) spheres which are free to move between retaining grids.

The mist elimination systems shown in Figures 1-1 and 1-2 each consists of a 3-pass, open-vane chevron mist eliminator with provision for underside and topside washing. During the early portion of the advanced test program, the TCA was also tested with a mist elimination system consisting of a wash tray (Koch Flexitray) in series with a 6-pass, closed-vane chevron mist eliminator, both with underside wash.

A typical system configuration used during venturi/spray tower lime reliability testing is shown in Figure 1-3. A typical system configuration used during TCA limestone utilization testing is shown in Figure 1-4. In the TCA configuration shown, three effluent hold tanks in series were used to increase limestone utilization.

This paper presents the results of mist elimination and limestone utilization testing at the Shawnee facility from June 1974 through January 1976. During this period the venturi/spray tower system was operated with both lime and limestone and the TCA system with limestone. All of the testing was performed under closed liquor loop conditions. Due to the relatively high inlet gas particulate loading, the slurry solids contained about 40 to 50 wt % fly ash.

For all of the tests during the reporting period, the sulfate (gypsum) saturation of the scrubber inlet liquor was maintained below about 140 percent and there was no significant accumulation of gypsum scale on scrubber internals. Previous testing has shown that scrubber internals can be kept relatively free of gypsum scale if the sulfate

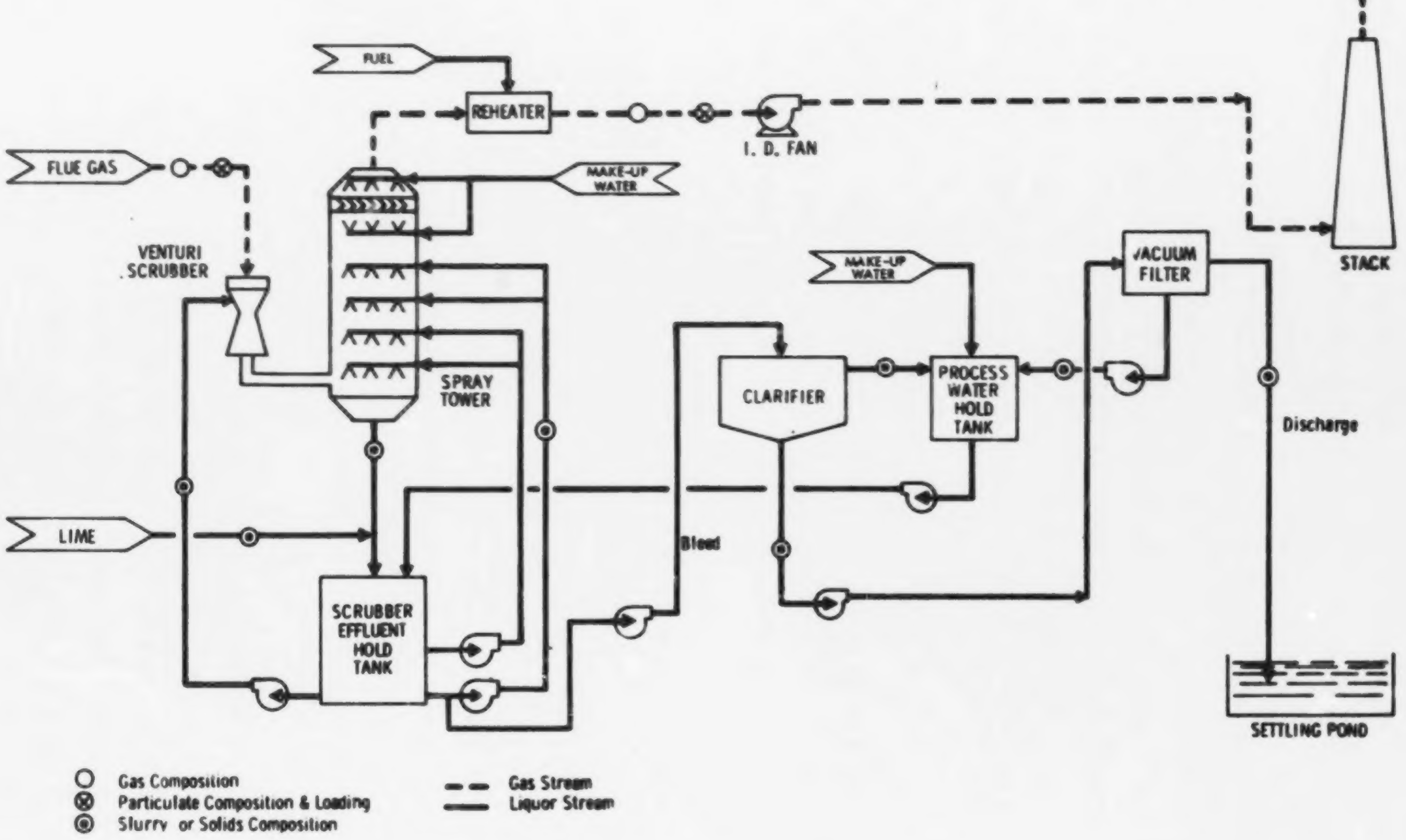

FUEL
REHEATER
I. D. FAN
STACK
FLUE GAS
MAKE-UP WATER
VENTURI SCRUBBER
SPRAY TOWER
MAKE-UP WATER
VACUUM FILTER
CLARIFIER
PROCESS WATER HOLD TANK
Discharge
LIME
Bleed
SCRUBBER EFFLUENT HOLD TANK
SETTLING POND
Gas Composition
Particulate Composition & Loading
Slurry or Solids Composition
Gas Stream
Liquor Stream

Figure 1-3. Typical Process Flow Diagram for Venturi/Spray Tower System

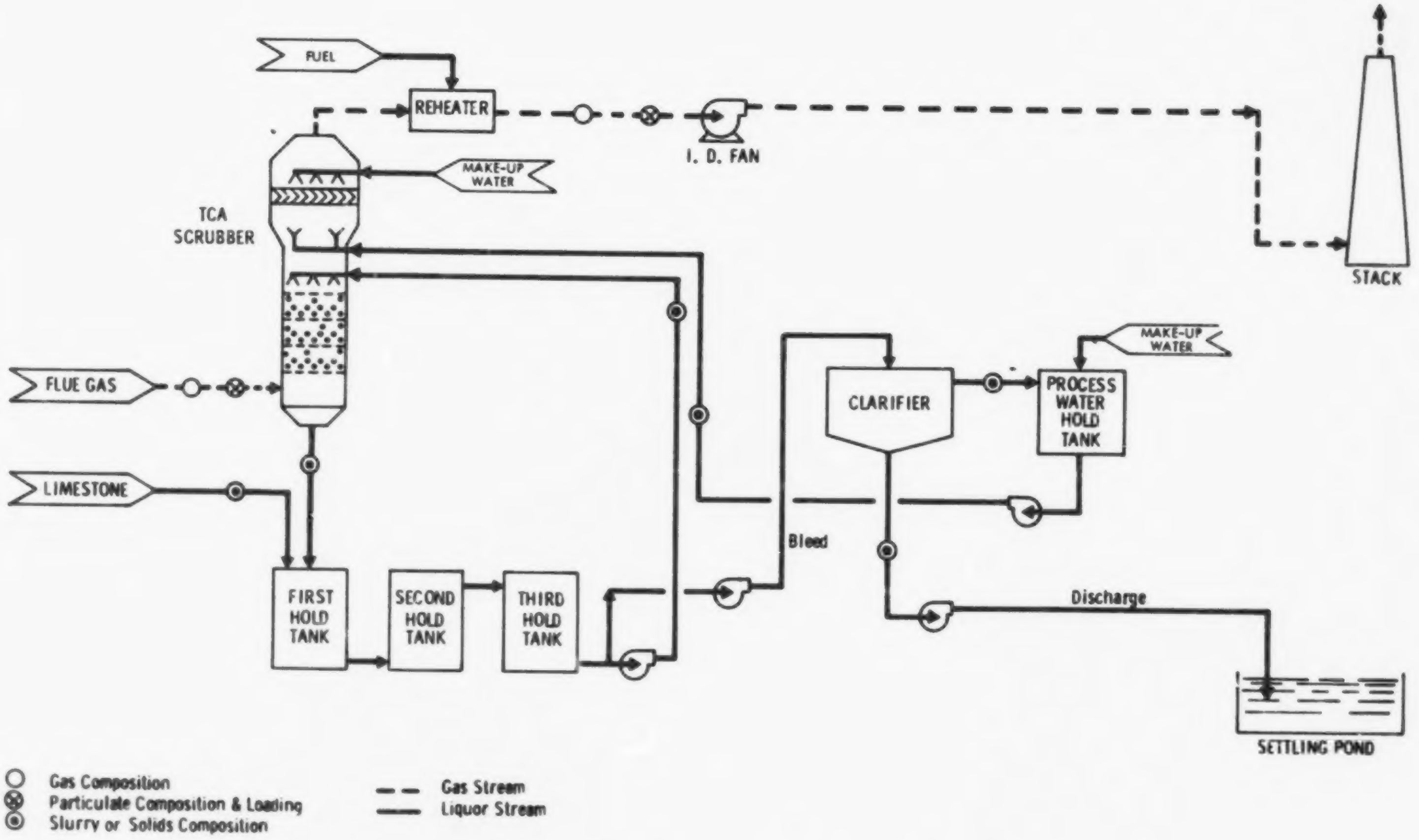

FUEL
REHEATER
I. D. FAN
STACK
MAKE-UP WATER
TCA SCRUBBER
FLUE GAS
LIMESTONE
FIRST HOLD TANK
SECOND HOLD TANK
THIRD HOLD TANK
Bleed
CLARIFIER
MAKE-UP WATER
PROCESS WATER HOLD TANK
Discharge
SETTLING POND
○ Gas Composition
⊗ Particulate Composition & Loading
⊙ Slurry or Solids Composition
– – Gas Stream
—— Liquor Stream

Figure 1-4. Typical Process Flow Diagram for TCA System

saturation of the scrubber liquor is kept below about 135 percent (see Section 7.3 in Reference 1). The effects of the process variables on scrubber liquor sulfate saturation observed during the advanced test program will be presented in a future paper.

Section 2

ADVANCED TEST PROGRAM OBJECTIVES AND SCHEDULE

The Shawnee Advanced Test Program is scheduled to run from June 1974 through June 1977. The objectives of the advanced program are:

- To continue long-term testing with emphasis on demonstrating reliable operation of the mist elimination systems.
- To investigate advanced process and equipment design variations for improving system reliability and economics. For example, testing will be conducted to investigate the practical upper limit of the gas velocity (i. e., minimum scrubber size) at which the scrubber mist elimination systems can be reliably operated. Also, tests will be conducted to evaluate system performance under conditions of minimum energy consumption for the desired levels of SO_2 and particulate removal.
- To evaluate process variations for substantially increasing limestone utilization and reducing sludge production. Tests will be conducted with scrubber effluent passing through three stirred tanks in series to approach a "plug-flow" condition and at reduced scrubber liquor pH to increase limestone utilization (see Reference 3).
- To evaluate scrubber operabiltiy during variable load (e. g., cycling gas rate) operation.
- To perform long-term (2 to 5 month) reliability testing on advanced process and equipment design variations.

- To investigate methods of improving waste solids separation. This may include testing a multiple-plate thickener, using coagulants, attempting to relate sludge characteristics to operating conditions, and making operational improvements on the centrifuge and filter.

- To study schemes for oxidizing sludge in order to improve solids settling characteristics and to reduce the chemical oxygen demand (COD) of the sludge. Oxidation tests will be conducted using a two-stage scrubber system developed by Borgwardt (Reference 4).

- To evaluate the effectiveness of three commercially offered sludge fixation processes and of untreated sludge disposal. Fixed sludges (Chemfix, Dravo, and IUCS) and untreated lime and limestone sludges are being continuously monitored in ponds at the Shawnee site. Aerospace Corporation is the major contractor and test director for this effort.

- To evaluate system performance and reliability without fly ash in the flue gas. Tests will be conducted with flue gas taken downstream of the Shawnee boiler No. 10 electrostatic precipitator, i.e., with less than 0.1 grain/scf of particulate in the inlet flue gas.

- To determine the practical upper limits of SO_2 removal efficiency. Tests will be conducted to determine the practical upper limit of SO_2 removal by increasing the scrubber slurry pH, increasing the slurry rate, increasing the scrubber gas pressure drop, and adding magnesium ion (MgO) to the slurry.

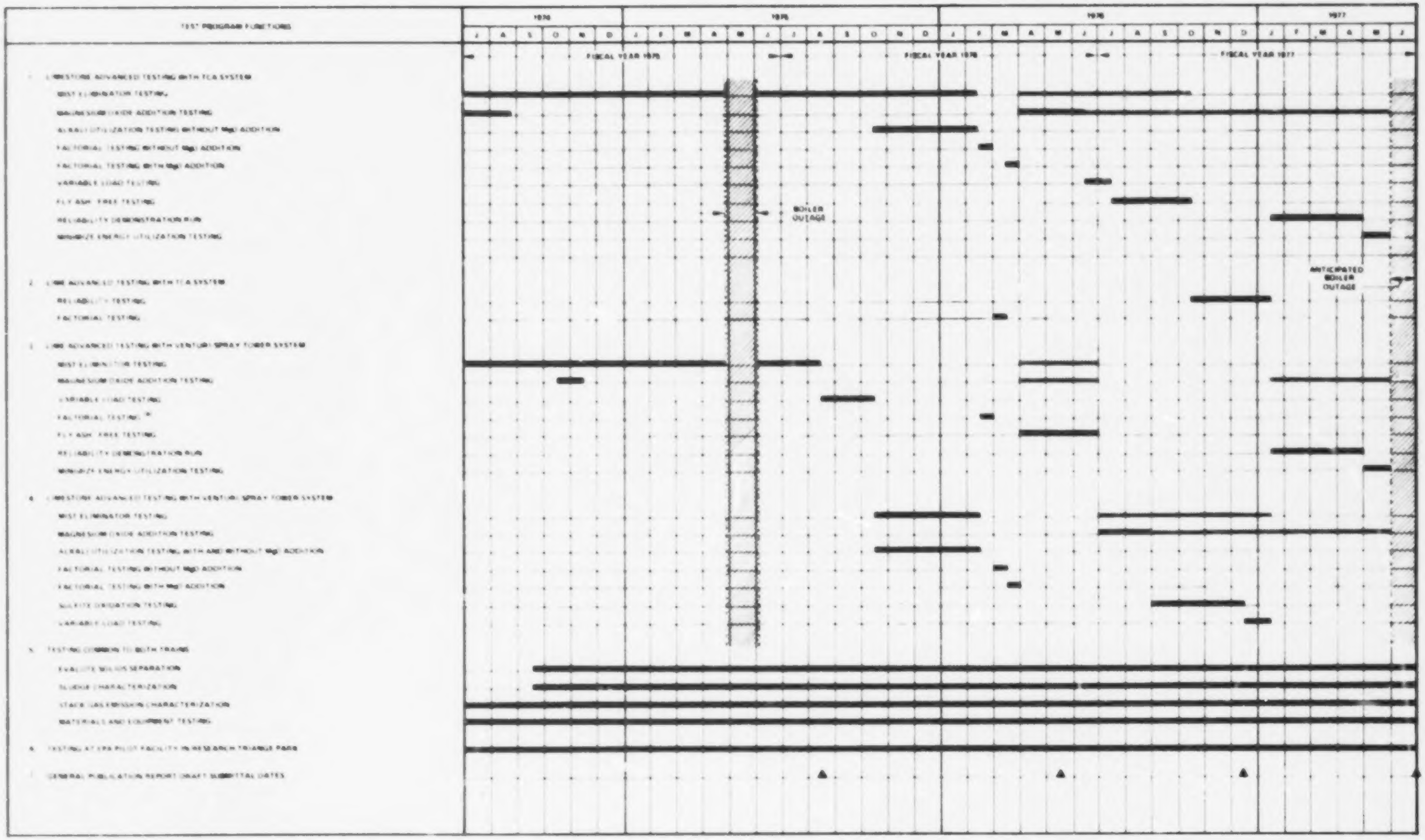
- To evaluate the TCA performance with lime and the venturi/spray tower performance with limestone.

- To characterize stack gas emissions including outlet particulate mass loading and size distribution, slurry entrainment, and total sulfate emission.

- To evaluate, under the direction of TVA, corrosion and wear of alternative plant equipment components and materials.

- To develop a computer program, in conjunction with TVA, for the design and cost comparison of full-scale lime and limestone systems.

The current test program schedule, based on the defined objectives, is presented in Figure 2-1. As can be seen in the figure, as of January 1976, limestone tests were in progress on both the venturi/ spray tower and TCA systems to: (1) demonstrate the reliability of the mist elimination systems and (2) determine the effects of process variables on limestone utilization.



(a) INCLUDES: MAXIMIZING SO_2 REMOVAL EFFICIENCY. TESTING MAY INCLUDE ADDITION OF MgO

NOTE: DASHED LINES REPRESENT SEQUENCES WHICH MAY BE CONTINUED DURING THE PERIOD INDICATED

Figure 2-1. Current Schedule for Shawnee Advanced Test Program

Section 3

LIME TESTING WITH A CHEVRON MIST ELIMINATOR IN THE VENTURI/SPRAY TOWER SYSTEM

This section summarizes the results of mist elimination reliability testing from June 1974 through mid-October 1975 with lime on the venturi/spray tower system. During this period, a 3-pass, open-vane, 316 stainless steel chevron mist eliminator was used in the spray tower.

3.1 TESTING AT CONSTANT GAS VELOCITY

During the period of advanced testing from June 1974 through March 1975, several mist eliminator washing configurations were evaluated under the following typical test conditions:

Spray tower gas velocity	6.7 ft/sec*
Venturi liquid-to-gas ratio	30 gal/mcf
Spray tower liquid-to-gas ratio	60 gal/mcf
Weight percent solids recirculated	8
Effluent residence time	12-24 minutes
Scrubber inlet slurry pH (controlled)	8
Weight percent solids in discharge cake	43-60
Inlet gas SO_2 concentration	1150-4250 ppm

---

*In this report, all gas velocities and liquid-to-gas ratios are at scrubber operating conditions, i.e., saturated gas at scrubber temperature. With flue gas operation, the scrubber temperature is approximately 125°F. The gas velocities are all superficial velocities.

For these test conditions, SO_2 removal was 75 to 95 percent, lime utilization[*] was approximately 90 percent, the scrubber outlet liquor pH was approximately 5.0, the sulfate (gypsum) saturation of the scrubber inlet liquor was 105 to 140 percent, the dissolved solids in the scrubber liquor was 6000 to 12000 ppm, and the total pressure drop was about 12.5 inches H_2O, including 9 inches H_2O across the venturi.

Initially (Runs 604-1A through 608-1A[**]) the mist eliminator was washed from the bottomside only. With this wash configuration, soft solids accumulation did not occur on the mist eliminator, but there was a continual problem of gypsum scale formation on the top vanes. The runs showed that this scale formation rate on the top mist eliminator vanes (25 to 50 mils/week) was relatively unaffected by the wash cycle, wash rate, or quality (sulfate saturation) of the wash liquor. Run 608-1A showed that intermittent washing with high pressure (45 psig) raw water at a rate of 3 gpm/ft^2 for 9 minutes every 4 hours gave results that were at least as good as results with continuous washing with relatively low pressure raw water at 0.3 gpm/ft^2 (Run 606-1A). This is significant in that intermittent washing may be required in closed liquor loop operation due to restrictions in the allowable raw water makeup to the scrubber system.

---

[*] Percent alkali utilization is defined as 100 x moles SO_2 absorbed per mole of Ca added. The reciprocal of the fraction alkali utilization is stoichiometric ratio, defined as moles Ca added per mole of SO_2 absorbed.

[**] Detailed operating conditions for Runs 604-1A through 624-1A can be found in References 1 and 2.

Starting in late September 1974, runs were made (Runs 609-1A and 610-1A) in which the entire mist eliminator underside and a 14 ft^2 area[*] on the topside were washed at high pressure (45 psig) with makeup water at a rate of 2.7 gpm/ft^2 for the underside and 1.0 gpm/ft^2 for the topside in an intermittent cycle of about 8 minutes every 4 hours. After 530 hours of operation, the washed area of the top mist eliminator vanes was essentially clean (less than 1 mil of solids accumulation) compared with an average of 70 mils scale buildup on the rest of the topside vane surfaces. The mist eliminator bottom vanes were covered with smooth white scale about 10 mils thick.

In the final run at 6.7 ft/sec spray tower gas velocity (Run 623-1A), begun in early March 1975, a wash system was installed in the spray tower to provide sequential and intermittent washing of the topside vanes of the chevron mist eliminator. The topside wash was accomplished by operating 6 nozzles in sequence. Every 80 minutes, one nozzle was actuated for 4 minutes at a rate of 0.5 gpm/ft^2 at 13 psig. The underside wash rate was 3 gpm/ft^2 at 45 psig for 3.5 minutes every 4 hours.

---

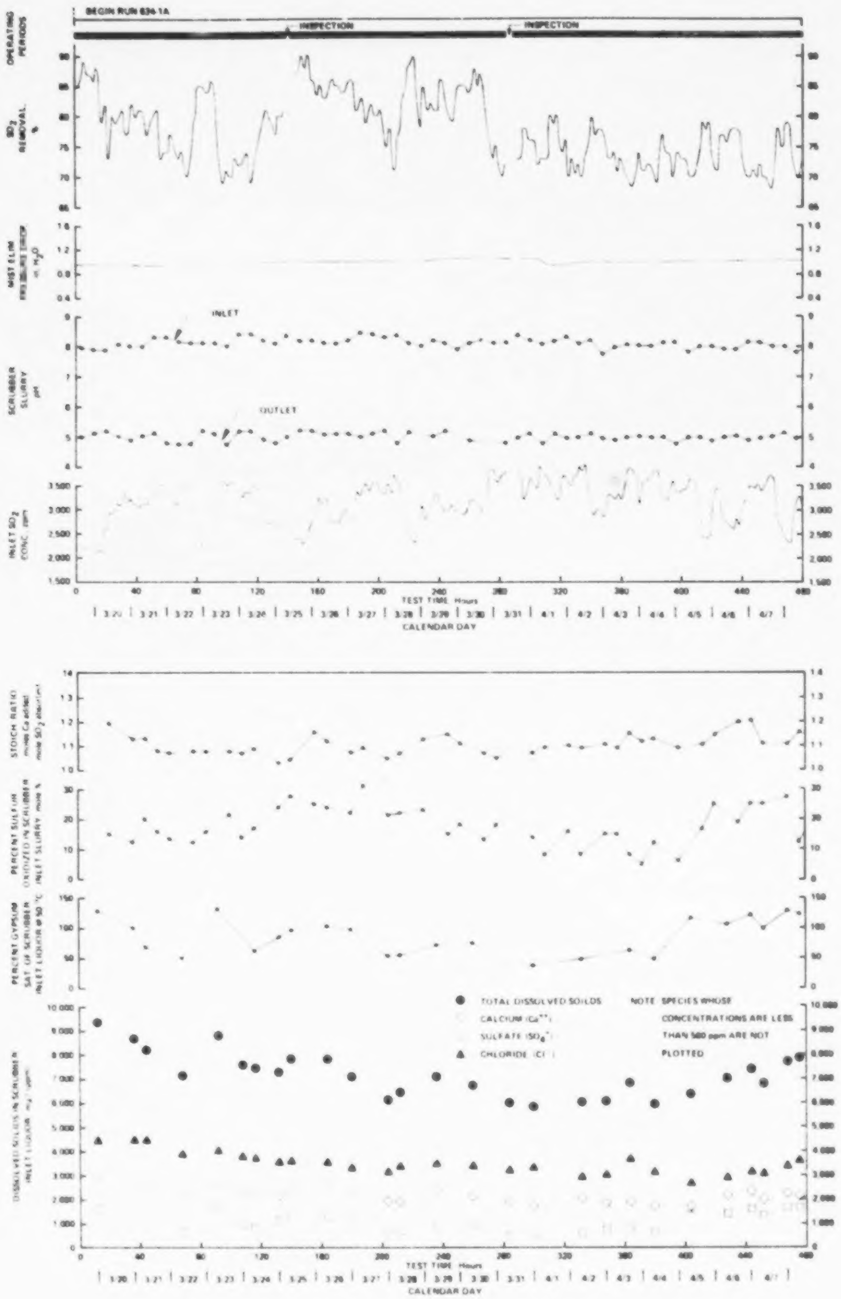
[*] Only a small section of the topside was washed because of a concern that entrainment from the top spray might overload the reheater and possibly allow moisture to reach the fan. In later tests, the entire topside was washed by resorting to sequential washing. Another possible solution would have been to use a second mist eliminator to intercept entrainment from the topside sprays.

After 162 hours of operation, an inspection showed that the combination of sequential topside wash and intermittent underside wash was successful in preventing scale and solids accumulation on the mist eliminator. Only scattered dust a few mils thick was formed on some of the vanes.

A new run (Run 624-1A) was started on March 19, 1975, without system cleaning, to test the mist elimination system at a higher spray tower gas velocity of 8.0 ft/sec. For this run, the intermittent wash cycle for the underside of the mist eliminator was increased from 3.5 to 4.3 minutes every 4 hours because of the higher makeup water rate available at the higher gas velocity. The wash rate and cycle for the topside were not changed from the previous run. Run 624-1A was terminated on April 23, 1975, after 823 hours of operation due to a scheduled 5-week maintenance outage on boiler No. 10.

A total of 4 inspections were made for this run, including one at the end of the run. The mist eliminator was found to be free of any restriction during each of these inspections. Vane surfaces washed earlier in the wash cycle usually held 2 mils of scattered white dust, while recently washed vane surfaces were entirely clean. Operating data for the initial 480 hours of Run 624-1A are presented in Figure 3-1.

Following the boiler outage, a run was begun on June 20, 1975, (Run 625-1A) to continue testing the mist eliminator at 8 ft/sec spray tower gas velocity with a reduced underside wash rate of 1.5 gpm/ft^2 for 4.3 minutes every 4 hours. The wash rate and cycle for the topside were the same as for the previous two runs. Only about one-half of the makeup



Gas Rate : 30,000 acfm @ 330 °F
Liquor Rate to Venturi : 600 gpm
Liquor Rate to Spray Tower : 1200 gpm
Venturi L/G : 25 gal/mcf
Spray Tower L/G : 50 gal/mcf
Spray Tower Gas Velocity : 8.0 ft/sec
No. of Spray Headers : 4
EHT Residence Time : 17 min

Percent Solids Recirculated : 7-10 wt %
Venturi Pressure Drop : 9 in H_2O
Total Pressure Drop, Excluding Mist Elim. : 12.5-12.6 in H_2O
Scrubber Inlet Liquor Temperature : 120-126 °F
Liquid Conductivity : 6,100-10,000 μ mhos/cm
Discharge (Clarifier and Filter) Solids
Concentration : 48-54 wt %
Lime Addition to Scrubber Downcomer

Figure 3-1. Operating Data for Venturi/Spray Tower Run 624-1A

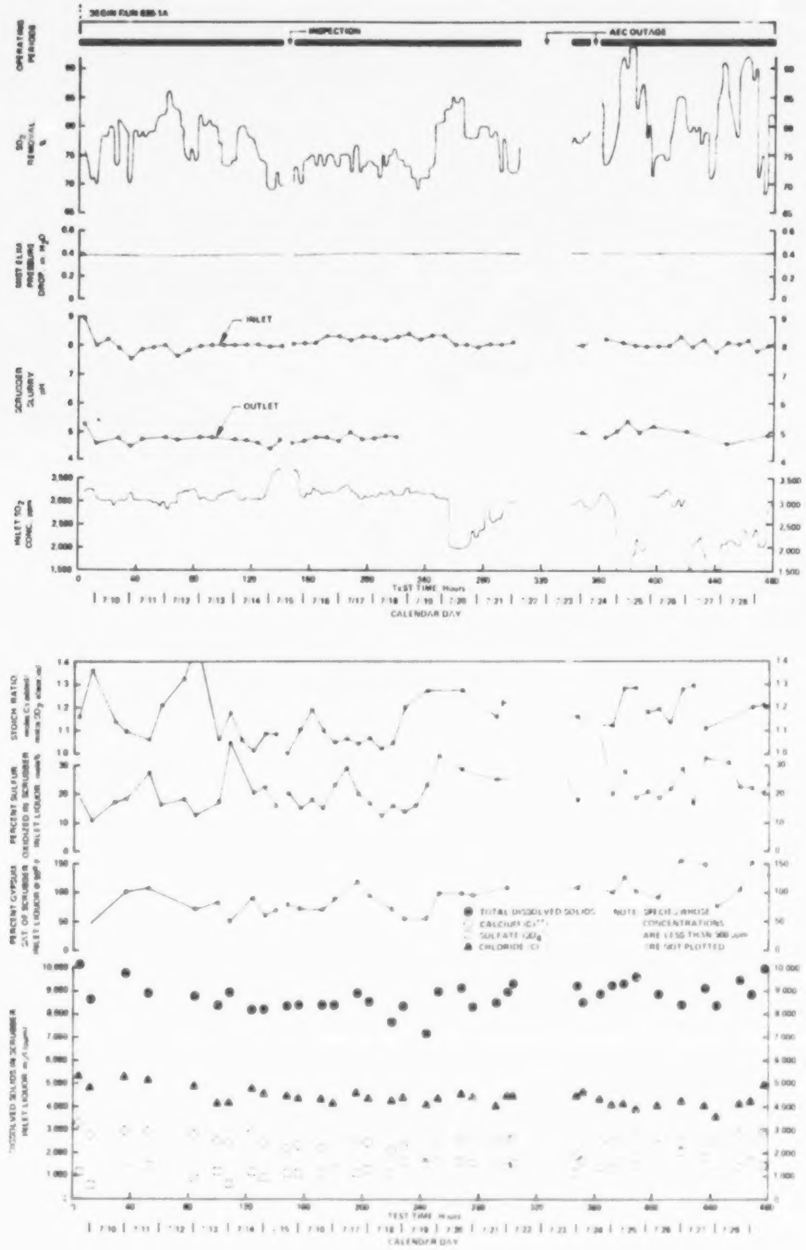
water available during closed liquor loop operation was needed to wash the mist eliminator. After 319 operating hours, the chevron mist eliminator was essentially clean (less than 2 percent restricted with dust).

The spray tower gas velocity was increased to the maximum achieveable value of 9.4 ft/sec in the next run (Run 626-1A) which started on July 9, 1975, with no cleaning of the mist eliminator. The intermittent wash cycle for the underside of the mist eliminator was increased to 6 minutes every 4 hours because of the higher makeup water rate available at the higher gas velocity. The topside wash rate and cycle were unchanged. As in the previous run, only about one-half of the makeup water available during closed liquor loop operation was needed to wash the mist eliminator. An inspection after 569 hours of operation showed that the chevron mist eliminator had remained essentially clean during the test (less than 2 percent restricted). Operating data for the initial 480 hours of Run 626-1A are presented in Figure 3-2.

The recirculation slurry solids concentration was increased from 8 wt % to 15 wt % in Run 627-1A which began on August 5, 1975. The spray tower gas velocity was maintained at 9.4 ft/sec. During this run, the inlet gas SO_2 concentration dropped to a low value of about 1500 ppm from August 6 through August 8.* As a consequence during that period, the outlet scrubber liquor pH rose to about 6.2 (the scrubber inlet liquor pH was controlled at 8.0). An inspection

---

*Low sulfur Montana coal was being burned in boiler No. 10.



Gas Rate = 35,000 acfm @ 330 °F
Spray Tower Gas Velocity = 8.4 ft/sec
Liquor Rate to Venturi = 600 gpm
Liquor Rate to Spray Tower = 1,400 gpm
Venturi L/G = 21 gal/mcf
Spray Tower L/G = 50 gal/mcf
No. of Spray Headers = 4
EHT Residence Time = 12 min

Percent Solids Recirculated = 8-9 wt %
Venturi Pressure Drop = 9 in. H_2O
Total Pressure Drop, Excluding Mist Elim. = 14.2-15 in. H_2O
Scrubber Inlet Liquor Temperature = 130-133 °F
Liquid Conductivity = 11,000-17,000 μ mhos/cm
Discharge (Clarifier and Filter) Solids
Concentration = 52-60 wt %
Lime Addition to Scrubber Downcomer

Figure 3-2. Operating Data for Venturi/Spray Tower Run 626-1A

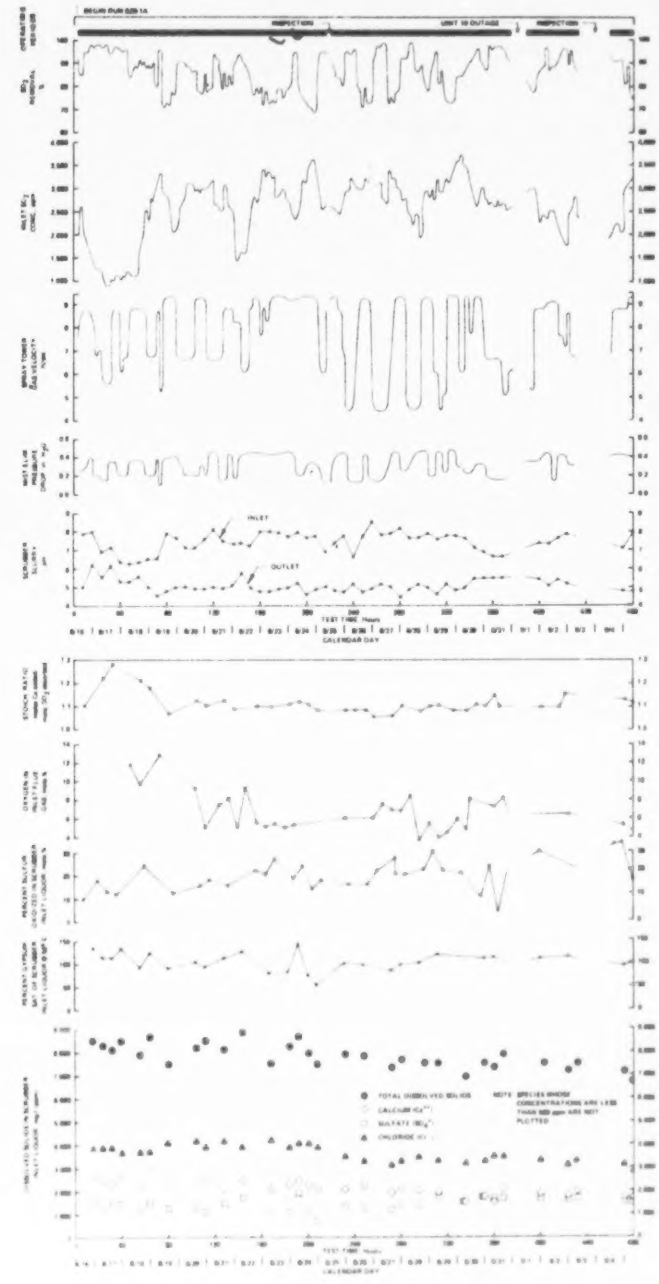
on August 8 showed that the higher level of liquor pH within the spray tower resulted in calcium sulfite scale formation within the tower and on the lower vanes of the chevron mist eliminator.* An inspection at the end of the run, after 187 hours of operation, showed that the scale had diminished somewhat during the latter half of the test, and that the mist eliminator was 2 to 3 percent restricted with sulfite scale and dust. At the conclusion of Run 627-1A, the mist eliminator had been in operation for 1075 hours without cleaning, including 319 hours at 8.0 ft/sec and 756 hours at 9.4 ft/sec gas velocity.

3.2 VARIABLE LOAD TESTING

Run 628-1A was begun on August 16 to test the operability and controllability of the venturi/spray tower system under cycling gas load. The gas flow rate for this run (17,000 to 35,000 acfm) was varied with the actual Unit No. 10 boiler load (about 60 to 160 Mw), resulting in spray tower gas velocities between 4.5 and 9.4 ft/sec. Constant liquor rates were used in the venturi and spray tower. The pressure drop across the venturi was held constant at 9 inches H_2O for all gas rates by varying the plug position. The same mist eliminator wash scheme as in Runs 626-1A and 627-1A was used during the test. Operating data for the initial 480 hours of Run 628-1A are presented in Figure 3-3. As can be seen in the figure, the inlet gas SO_2 concentration varied between 1500 and 4200 ppm during the test.

---

*This problem of sulfite scale formation at high outlet scrubber liquor pH was corrected in subsequent testing by controlling the outlet scrubber liquor pH to 5.0±0.5, with the constraint that the scrubber inlet liquor pH remain equal to or less than 8.0.



Gas Rate = 17,000-35,000 acfm @ 330 °F
Spray Tower Gas Velocity = 4.5-9.4 ft/sec
Liquor Rate to Venturi = 600 gpm
Liquor Rate to Spray Tower = 1,400, 1,600 (after 8/25)
Venturi L/G = 21-44 gal/mcf
Spray Tower L/G = 50-117 gal/mcf
No. of Spray Headers = 4
EHT Residence Time = 12 min

Percent Solids Recirculated = 8.9-11.4 wt %
Venturi Pressure Drop = 9 in. H_2O
Total Pressure Drop, Excluding Mist Elim. = 10.3-14.3 in. H_2O
Scrubber Inlet Liquor Temperature = 124-133 °F
Liquid Conductivity = 6,200-11,400 μ mhos/cm
Discharge (Clarifier and Filter) Solids
Concentration = 51-55 wt %
Lime Addition to Scrubbe Downcomer

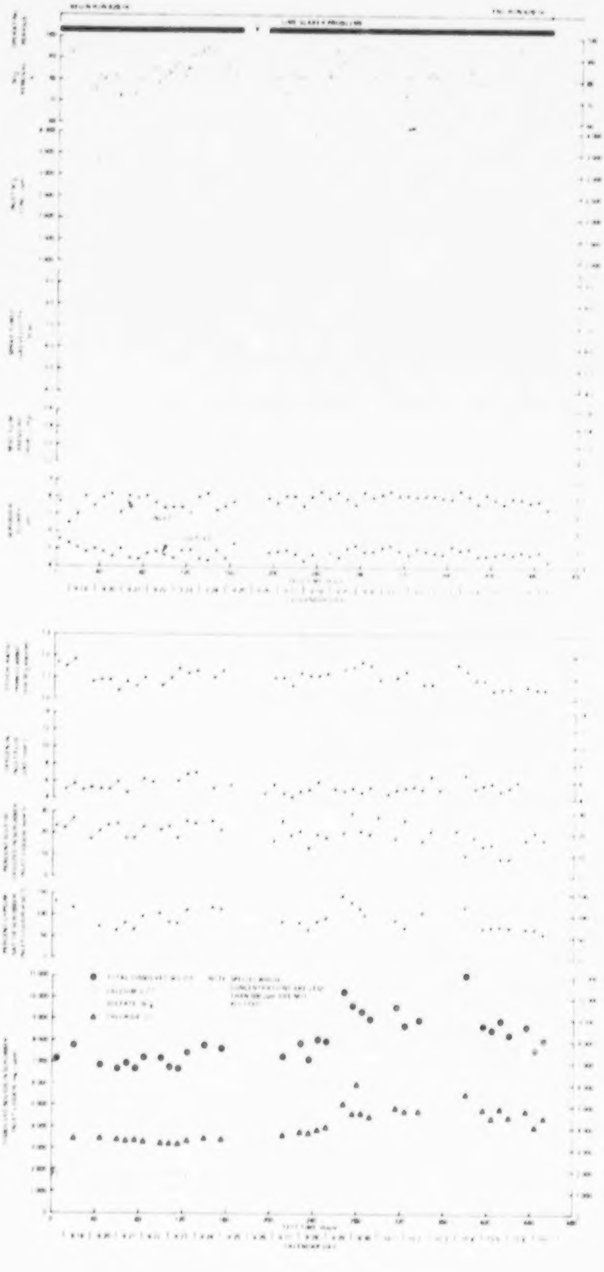
Figure 3-3. Operating Data for Venturi/Spray Tower Variable Load Run 628-1A

An inspection after 717 hours of operation showed that the mist eliminator was essentially clean (only 2 percent restricted with dust). No problems were experienced in controlling the system during the test.

Run 628-1B was a continuation of 628-1A, except that the venturi plug position was fixed to give a maximum venturi pressure drop of 9 inches H_2O at the maximum gas rate of 35,000 acfm. Operating data for this test are presented in Figure 3-4. An inspection after 426 hours of operation (for a total of 1143 hours for Runs 628-1A and 628-1B) showed that the mist eliminator was essentially unchanged (2 percent restricted with dust).

3.3 CONCLUSIONS

The long-term (6-12 month) operability of the 3-pass, open-vane chevron mist eliminator in the venturi/spray tower system with lime additive has been demonstrated for intermittent high pressure underside washing and sequential low pressure topside washing with raw water. The mist eliminator remained essentially free of solids deposit for 756 hours of operation at a spray tower gas velocity of 9.4 ft/sec and with 8 to 15 wt % solids in the recirculating slurry. The mist eliminator also remained free of solids deposit during variable load (cycling gas rate) tests lasting a total of 1143 hours. For the lime testing, only about one-half of the makeup water available during closed liquor loop operation was needed to wash the mist eliminator. The lime utilization for these tests was about 90 percent.



Reproduced from best available copy.

Gas Rate: 19,000-35,000 acfm @ 330 °F
Spray Tower Gas Velocity: 5.1-9.4 ft/sec
Liquor Rate to Venturi: 600 gpm
Liquor Rate to Spray Tower: 1,600 gpm
Venturi L/G: 21-39 gal/mcf
Spray Tower L/G: 57-105 gal/mcf
No. of Spray Headers: 4
EHT Residence Time: 12 min

Percent Solids Recirculated: 8.6-9.7 wt %
Venturi Pressure Drop: 4.0-9.0 in. H_2O
Total Pressure Drop, Excluding Mist Elim.: 6.5-14.6 in. H_2O
Scrubber Inlet Liquor Temperature: 129-131 °F
Liquid Conductivity: 8,000-11,500 μ mhos/cm
Discharge (Clarifier and Filter) Solids
Concentration: 52-56 wt %
Lime Addition to Scrubber Downcomer

Figure 3-4. Operating Data from Venturi/Spray Tower Variable Load Run 628-1B

Section 4

LIMESTONE TESTING WITH A WASH TRAY AND CHEVRON MIST ELIMINATOR IN SERIES IN THE TCA SYSTEM

This section summarizes the results of mist eliminator reliability testing from September 1974 through March 1975 with limestone in the TCA system. During this period the mist elimination system consisted of a wash tray (Koch Flexitray) in series with a 6-pass, closed-vane, 316 stainless steel chevron mist eliminator.

4.1 TESTING AT 8.6 FT/SEC SUPERFICIAL GAS VELOCITY

TCA limestone Run 535-2A was begun on September 12, 1974, and continued through December 4, 1974, for a total of 1835 operating hours.* The TCA internals consisted of 3 beds (4 grids), with 5 inches per bed of 1 1/2 inch diameter, 6 gram TPR (thermoplastic rubber) spheres. The major test conditions at the start of the run were:

Gas velocity	8.6 ft/sec
Liquid-to-gas ratio	73 gal/mcf
Percent solids recirculated	15
Effluent residence time	12 min
Percent SO_2 removal (controlled)	84
Inlet gas SO_2 concentration	2000-4000 ppm
Weight percent solids in discharge sludge	35-42

---

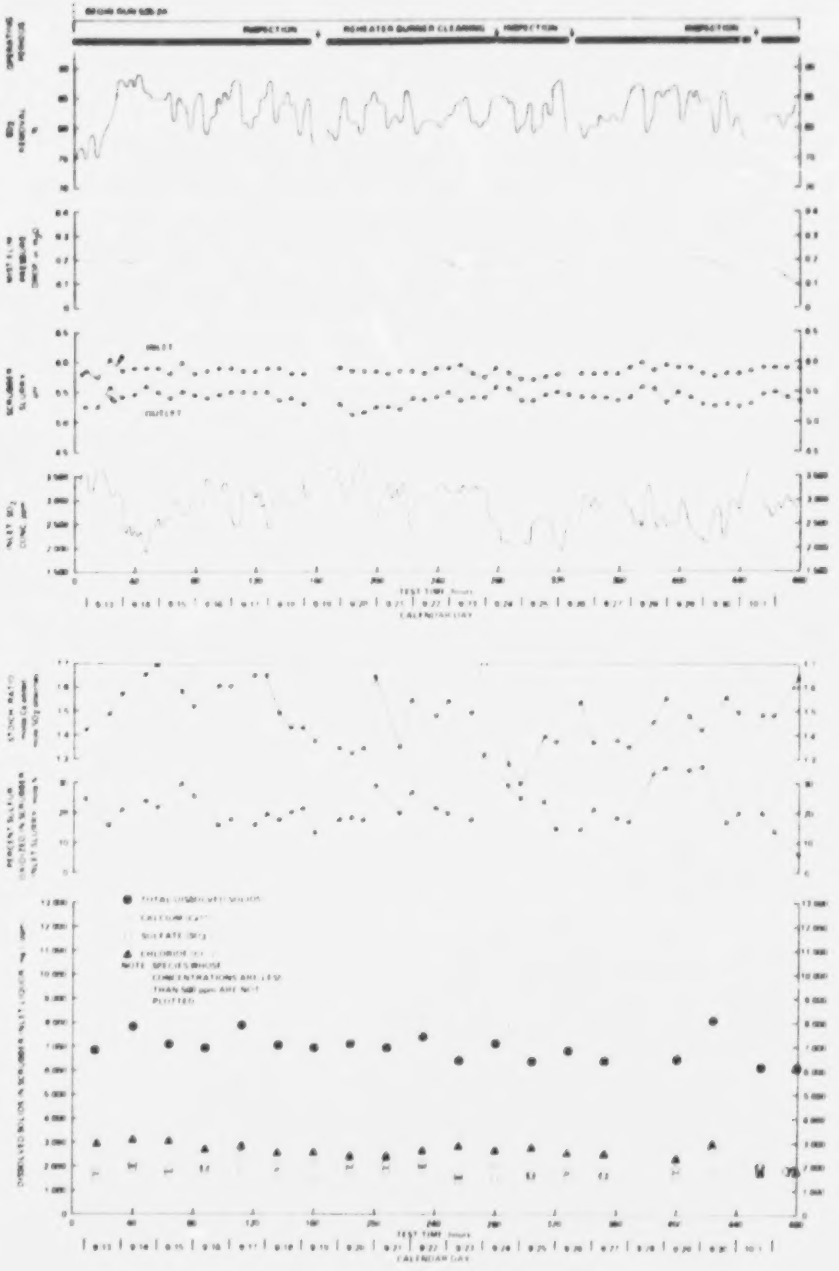
*TCA Run 535-2A is described in detail in Reference 2.

For these test conditions, the scrubber inlet liquor pH varied between 5.7 and 6.0, limestone utilization was approximately 65 percent (stoichiometric ratio of 1.54 moles Ca added/mole SO_2 absorbed), the sulfate (gypsum) saturation of the scrubber inlet liquor averaged 110 percent, the dissolved solids concentration in the scrubber liquor was 4000 to 8000 ppm, and the total pressure drop (including the mist elimination system) was 6 to 7 inches H_2O. Operating data for the initial 480 hours of testing are presented in Figure 4 1.

The clarified liquor return flow rate was maintained at a minimum of 15 gpm for Koch tray feed and mist eliminator wash to prevent high sulfate supersaturation in the wash tray effluent liquor. The underside of the mist eliminator was washed continuously with 15 gpm (0.31 gpm/ft^2) diluted clarified liquor (about 9 gpm makeup water plus about 6 gpm clarified liquor). The tray was fed with the 15 gpm mist eliminator wash plus the remaining ~9 gpm (minimum) of clarified liquor (0.50 gpm/ft^2). The underside of the tray was sparged with 125 psig steam for one minute each hour.

Throughout the first half of Run 535-2A, there was a continual problem of solids accumulation on the scrubber walls between the steam sparger and the main slurry spray header. This problem was resolved by installing four wall spray nozzles utilizing the wash tray effluent liquor. It was observed that the wall sprays were also effective in washing the underside of the tray where it was covered by the sprays.

After 1835 operating hours for Run 535-2A, an inspection showed that the top surface of the wash tray was entirely clean and that the chevron



Gas Rate = 20,500 acfm @ 300 °F
Liquor Rate = 1200 gpm
L/G = 73 gal/mcf
Gas Velocity = 8.6 ft/sec
EHT (Sealed) Residence Time = 12 min (8/12-9/27)
15 min (after 9/27)
Three Stages, 5 in spheres/stage

Percent Solids Recirculated = 12-15 wt %
Total Pressure Drop, Excluding Mist Elim.
and Koch Tray = 4.0-4.6 in H_2O
Scrubber Inlet Liquor Temperature = 120-126 °F
Liquid Conductivity = 4,800-10,000 μ mhos/cm
Discharge (Clarifier) Solids
Concentration = 35-42 wt %

Figure 4-1. Operating Data for TCA Run 535-2A

mist eliminator was less than 5 percent restricted by solids (mostly fly ash). The underside of the wash tray had no solids accumulations greater than 1/4 inch thick. These deposits did not interfere with tray operation. Also, the scrubber walls between the tray and slurry spray header were clean.

Run 535-2A was arbitrarily terminated on December 4, 1974, and Run 535-2B begun, when low sulfur Montana coal was unexpectedly introduced into Shawnee boiler No. 10. The wide swings in inlet gas SO_2 concentration (1200 to 4900 ppm in 8 hours) caused system upsets which resulted in some scale formation on the wash tray and solids accumulation in the chevron mist eliminator. During Run 535-2B, the steam sparger was removed* and the four wall spray nozzles were replaced by a single tray under wash nozzle, which also provided rinsing of the walls below the tray. This modification (see Figure 3-2 in Reference 2) proved to be effective in keeping both the underside of the wash tray and the walls below it clean.

4.2 TESTING AT 10 AND 12 FT/SEC SUPERFICIAL GAS VELOCITY

At scrubber gas velocities of 10 ft/sec or higher, plugging of the chevron mist eliminator by solids (mostly fly ash) became a problem. At 10 ft/sec scrubber gas velocity (6.5 ft/sec in the mist eliminator and wash tray areas), the overall restriction of the mist eliminator increased gradually to about 8 percent during the first 500 hours of operation and appeared to level out at 6 to 8 percent after 562 operating hours (Run 538-2A). At 12 ft/sec scrubber gas velocity (7.8 ft/sec in the mist eliminator and wash tray areas), the mist eliminator was 11 percent restricted by solids within only 215 hours of operation (Run 539-2A).

---

* In a full scale gas scrubbing unit, steam cleaning of the underside of a wash tray would be an economic burden.

4.3 CONCLUSIONS

The long-term operability of the TCA mist elimination system was demonstrated in an 1835-hour run in limestone service at 8.6 ft/sec superficial gas velocity and 15 wt % solids in the recirculating slurry. The mist elimination system consisted of a Koch Flexitray in series with a 6-pass, closed-vane, chevron mist eliminator, both with underside wash. Long-term operability was not demonstrated at increased gas velocity. At 10 ft/sec the chevron mist eliminator was 8 percent restricted with soft solids after 562 operating hours, while at 12 ft/sec the chevron mist eliminator was 11 percent restricted in only 215 hours.

Section 5

LIMESTONE UTILIZATION TESTING IN THE VENTURI/SPRAY TOWER AND TCA SYSTEMS

The results of limestone utilization testing from October 1975 through January 1976 on both the venturi/spray tower and the TCA systems are presented in this section. These tests were made as a result of a TVA economic study which showed that a potential existed for significantly improving the economics of limestone scrubbing by improving the utilization of the limestone feed. Improved limestone utilization not only results in a decrease in limestone feed requirements but also a corresponding decrease in waste sludge production.

Tests were conducted primarily to determine the effect of scrubber inlet liquor pH, effluent residence time, and hold tank design on limestone utilization. These variables were correlated with stoichiometric ratio (the reciprocal of fraction alkali utilization). For the venturi/spray tower system a single backmix effluent hold tank was used. For the TCA system both a single backmix hold tank and three backmix hold tanks in series (to simulate a plug flow reactor) were used.

For each combination of residence time and hold tank design, tests were conducted to cover a range of values of scrubber inlet liquor pH. Normally, the systems were run for about 4 to 5 days at a specified level of pH. During testing, the stoichiometric ratios were determined every 4 hours from solids analyses of the scrubber recirculation slurry.

The limestone used during the advanced program was "Fredonia Valley White", purchased from Fredonia Quarries, Fredonia, Kentucky. The limestone analysis was: 96 wt % $CaCO_3$, 1 wt % $MgCO_3$, and 3 wt % inerts. The limestone size distribution was: 90 wt % less than 325 mesh, 87 wt % less than 30 microns, 82 wt % less than 22 microns, and 52 wt % less than 6 microns.

Results of analyses presented here should be considered preliminary as the data reduction is still in progress. For example, no attempt has been made in this report to adjust the pH for chloride concentration in the slurry liquor, which ranged from 1500 to 6500 ppm during the testing.

Both the venturi/spray tower system and the TCA system were operated with a single stage, 3-pass, open-vane, 316 stainless steel mist eliminator with top and bottom wash. Details of the mist eliminator systems and the effect of alkali utilization on mist eliminator operation are discussed in Section 6.

5.1 UTILIZATION TESTING IN THE VENTURI/SPRAY TOWER SYSTEM WITH VARIABLE RESIDENCE TIME

In the venturi/spray tower system, limestone tests were made with a single backmix effluent hold tank to determine the effect of scrubber inlet liquor pH and residence time on stoichiometric ratio. Runs were made at 20, 12, and 6 minutes residence time. Major test conditions maintained constant during this period were:

Spray tower gas velocity	9.4 ft/sec
Venturi liquid-to-gas ratio	21 gal/mcf
Spray tower liquid-to-gas ratio	50-57 gal/mcf[*]
Venturi pressure drop	9 in. H_2O
Percent solids recirculated	15

Data showing the relationship between stoichiometric ratio and scrubber inlet liquor pH for the venturi/spray tower system are plotted in Figures 5-1, 5-2, and 5-3 for 20, 12, and 6 minutes residence time, respectively. As would be expected, scatter in the data was greatest at the shortest residence time, where pH recovery time is limited and a small variation in tank level can result in a significant change in residence time.

Sight average curves drawn through the data in Figures 5-1 through 5-3 are cross plotted in Figure 5-4, which shows the effect of effluent hold tank residence time and scrubber inlet liquor pH on stoichiometric ratio. Comparing 20 minutes residence time versus 12 in Figure 5-4, little effect of residence time on stoichiometric ratio can be seen. However, between 12 minutes and 6 minutes, a significant increase in stoichiometric ratio with decrease in residence time can be seen for scrubber inlet liquor pH's greater than or equal to about 5.6.

---

[*] All but 2 tests were at 50 gal/mcf.

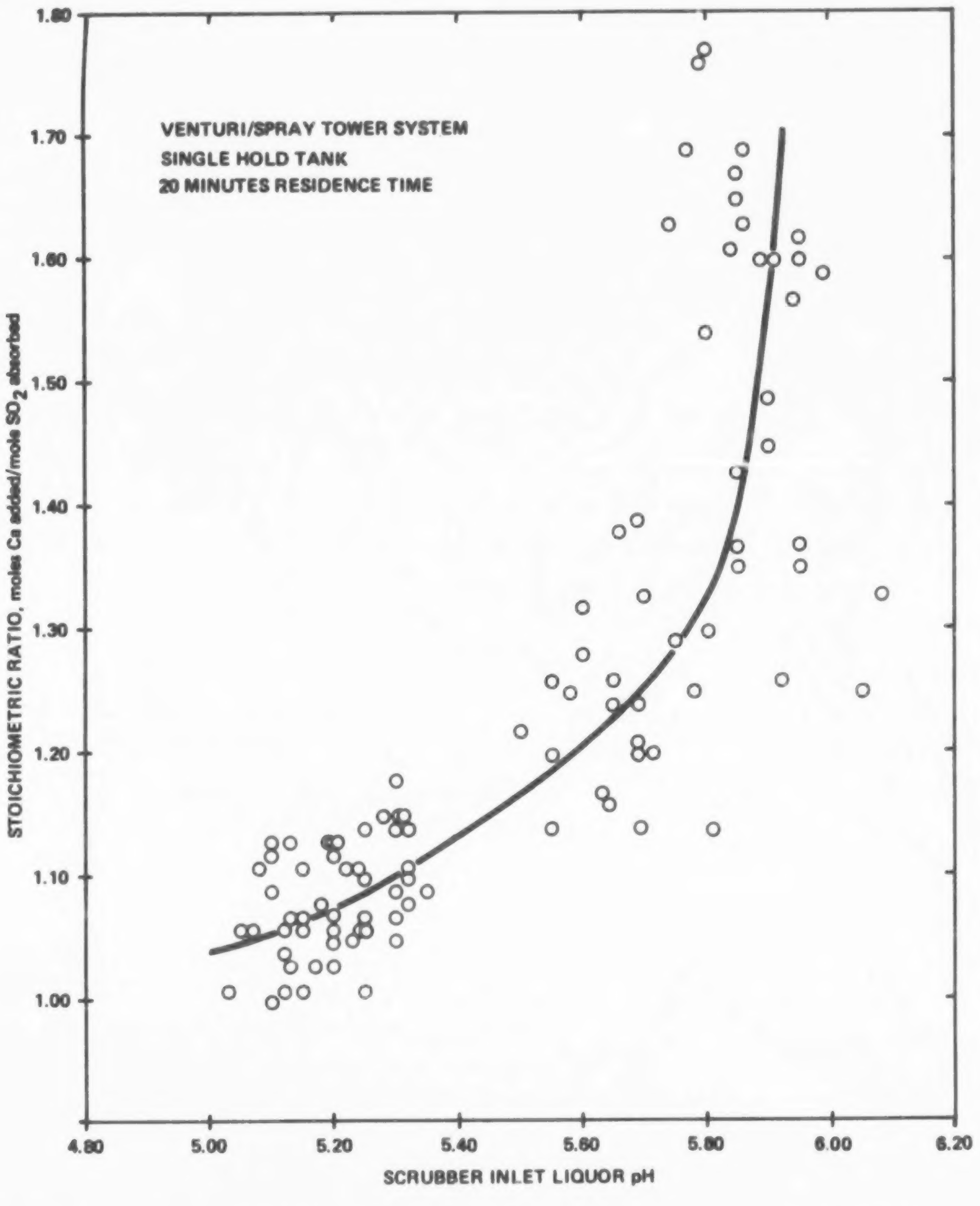


Figure 5-1. Stoichiometric Ratio versus Scrubber Inlet Liquor pH for a Single Hold Tank at 20 minutes Residence Time

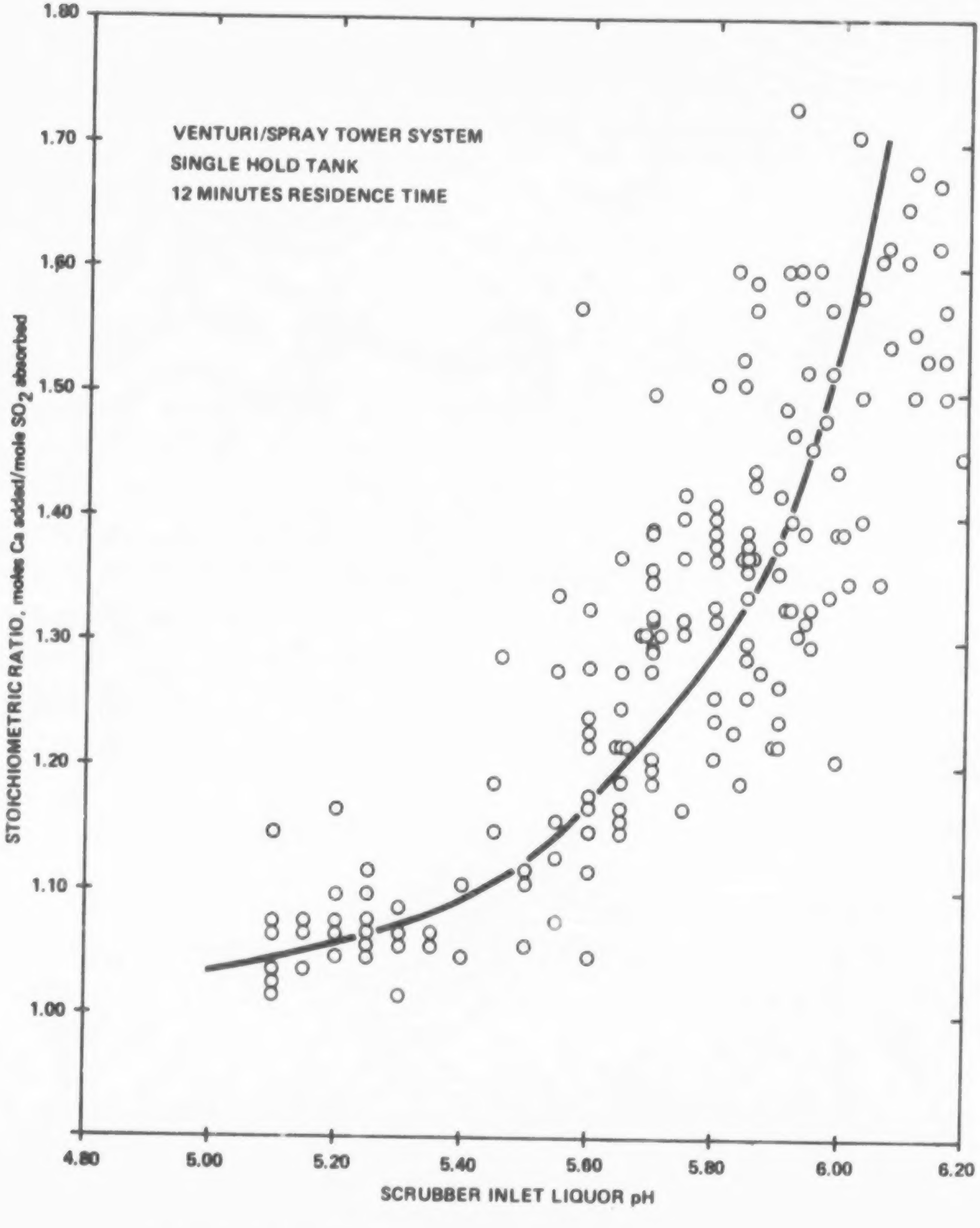


Figure 5-2. Stoichiometric Ratio versus Scrubber Inlet Liquor pH for a Single Hold Tank at 12 minutes Residence Time

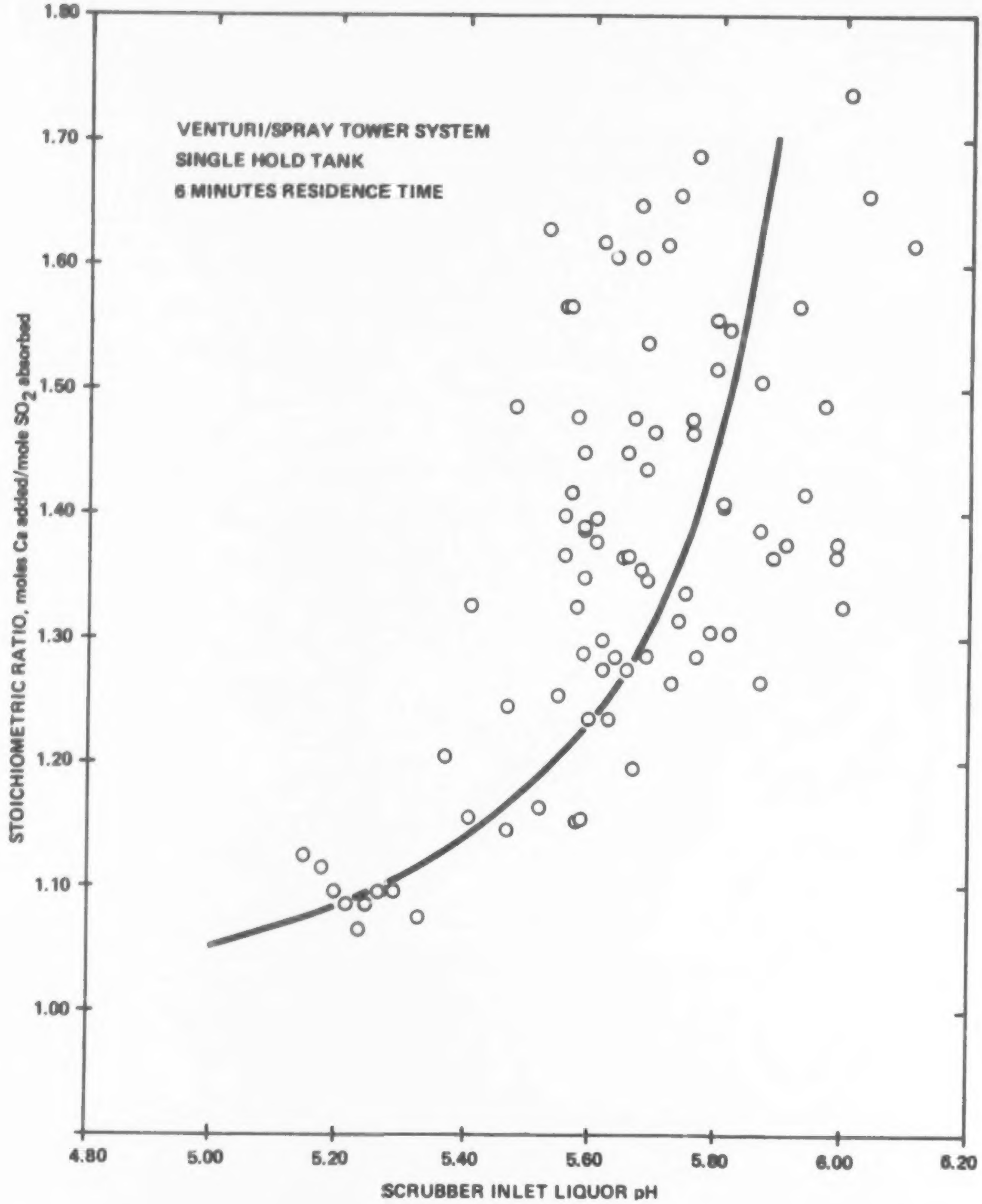


Figure 5-3. Stoichiometric Ratio versus Scrubber Inlet Liquor pH for a Single Hold Tank at 6 minutes Residence Time

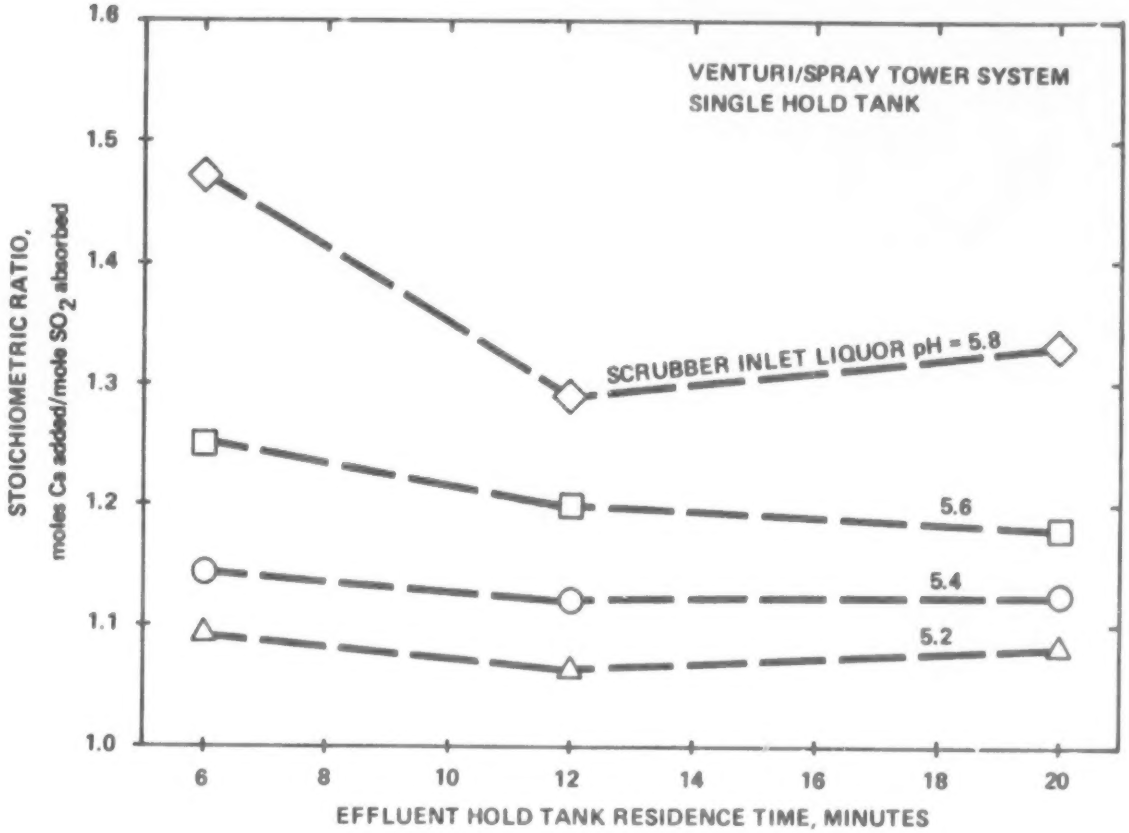


Figure 5-4. The Effect of Effluent Residence Time and Scrubber Inlet Liquor pH on Stoichiometric Ratio

Figure 5-5 shows the general relationship between SO_2 removal and stoichiometric ratio for an inlet gas SO_2 concentration range between 2500 and 3500 ppm. Data points have been plotted for hold tank residence times of 20, 12, and 6 minutes and a sight average line through all the data has been drawn. Referring back to Figure 5-4, it was shown that, for a given stoichiometric ratio greater than about 1.25, the scrubber inlet liquor pH is lower at 6 minutes residence time than at 12 or 20 minutes. Therefore, a corresponding reduction in SO_2 removal at 6 minutes residence time was expected. However, within the scatter of the data, such a decrease could not be discerned in Figure 5-5.

Figure 5-6 shows the relationship between SO_2 removal and scrubber inlet liquor pH for an inlet gas SO_2 concentration range from 2500 to 3500 ppm and a 12 minute hold tank residence time. Averages from Figure 5-6 and from similar plots for higher and lower inlet gas SO_2 concentration ranges at 12 minutes residence time are drawn in Figure 5-7, which shows the effect of inlet gas SO_2 concentration and scrubber inlet liquor pH on SO_2 removal. As expected, an increase in inlet SO_2 concentration results in a decrease in SO_2 removal at constant inlet pH (see Equlation 14-7 in Reference 1).

5.2 UTILIZATION TESTING IN THE TCA SYSTEM WITH THREE HOLD TANKS IN SERIES

Kinetic theory shows that for a continuous system where the reaction order is greater than zero, raw materials are more completely converted in a plug flow reactor than in a backmixed reactor of the same residence time. This concept for improving utilization was success-

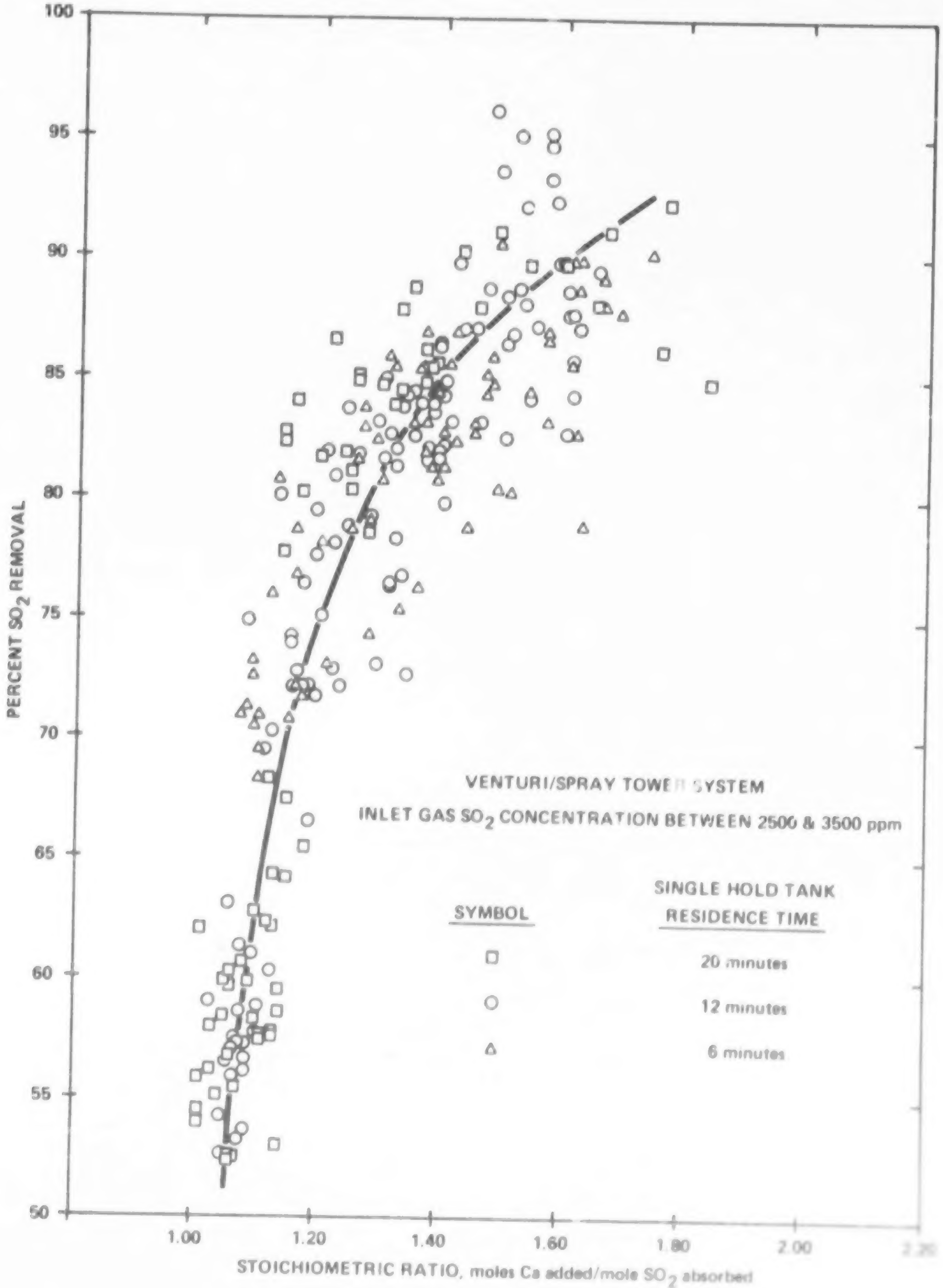


Figure 5-5. The Effect of Stoichiometric Ratio and Effluent Residence Time on Percent SO_2 Removal with a Single Hold Tank

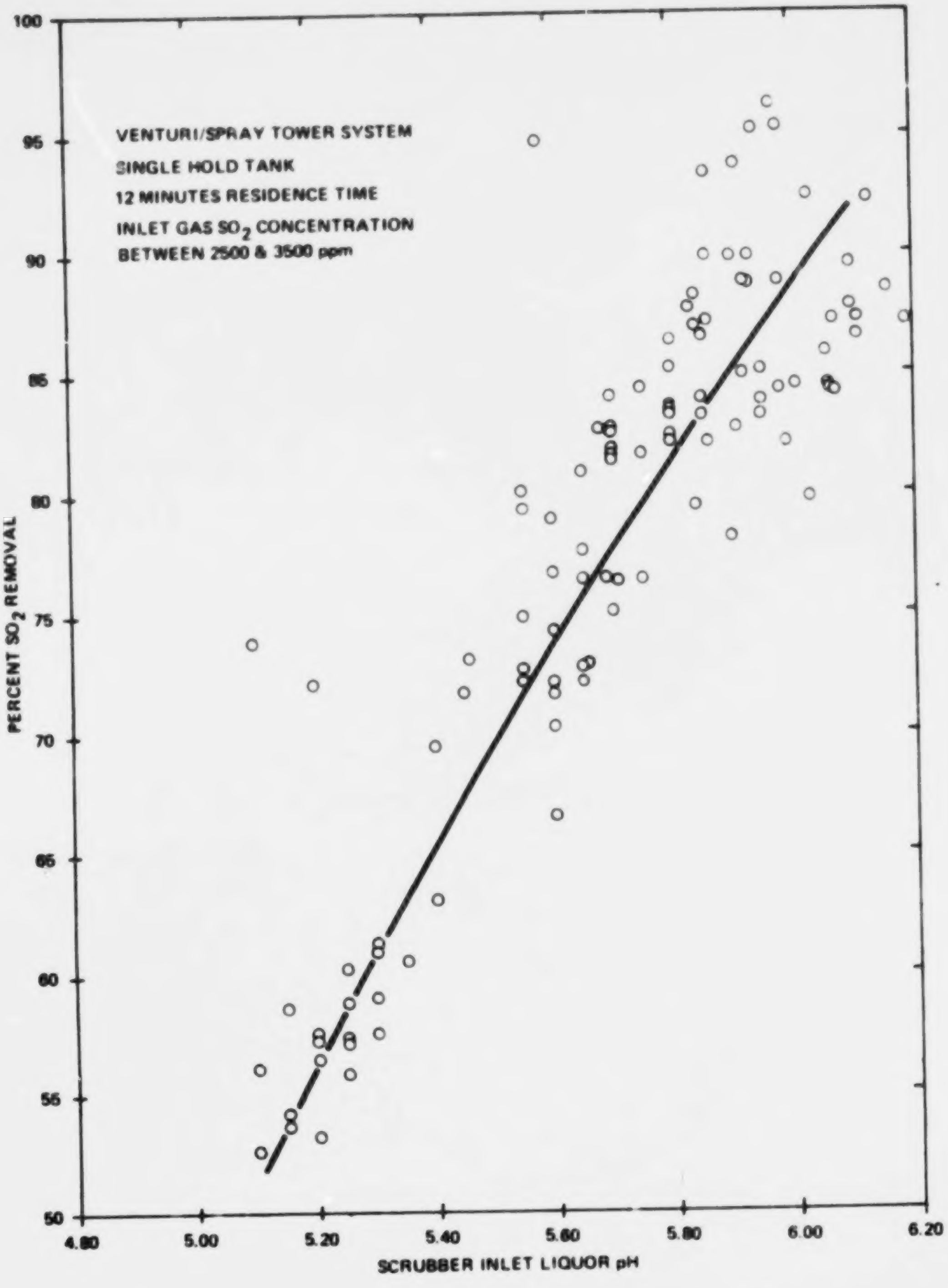


Figure 5-6. The Effect of Scrubber Inlet Liquor pH on Percent SO_2 Removal in the Venturi/Spray Tower System

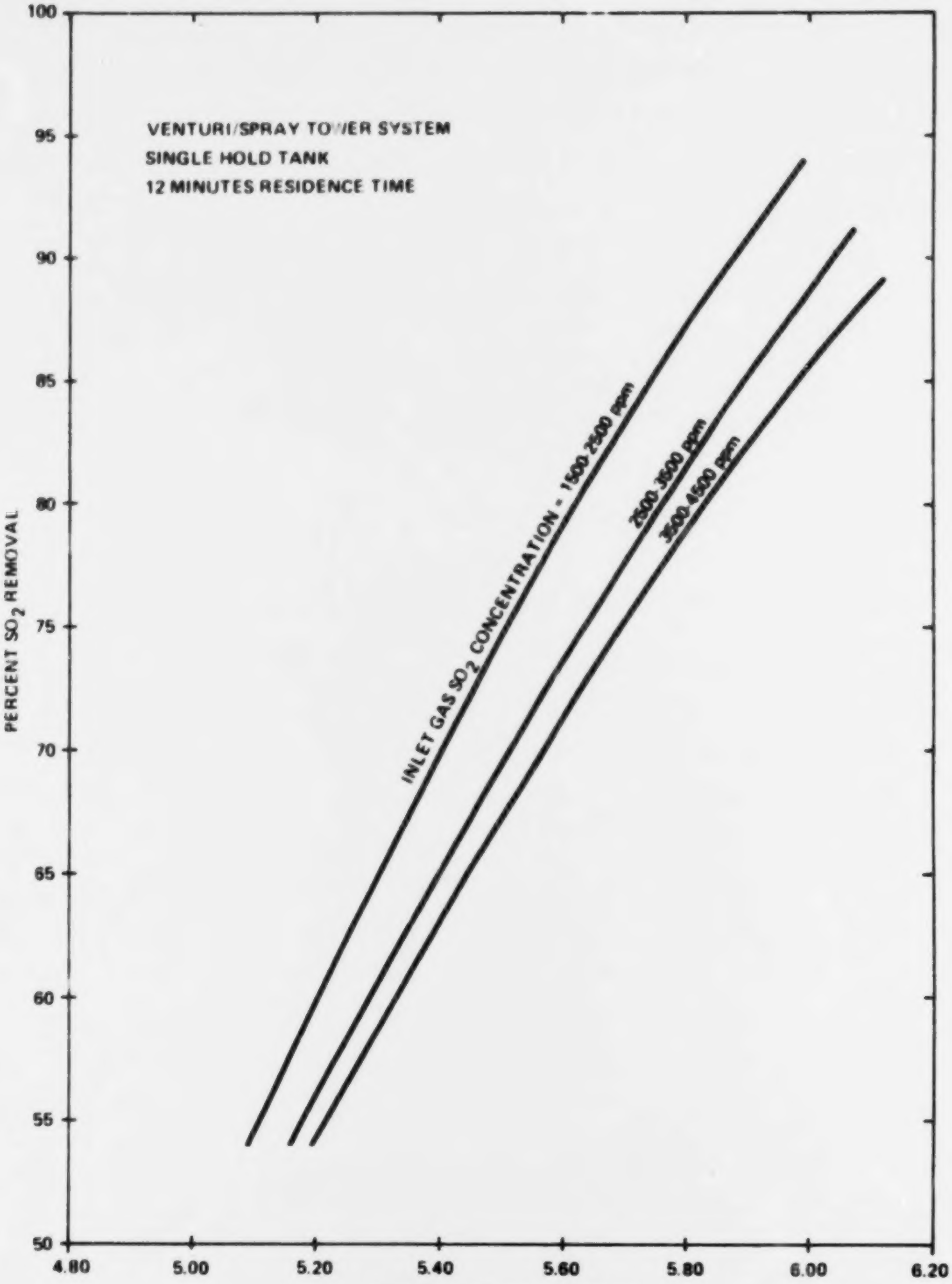


Figure 5-7. The Effect of Scrubber Inlet Liquor pH and Inlet Gas SO_2 Concentration on Percent SO_2 Removal

fully tested by Borgwardt (Reference 3) in a 0.1 Mw limestone scrubber with both a plug flow reactor and with 3 stirred tanks in series to approximate plug flow. The concept has now been tested with limestone in the TCA system at the Shawnee Test Facility (see Figure 1-4). Major conditions for these tests were:

TCA superficial gas velocity	12.5 ft/sec
Liquid-to-gas ratio	50 gal/mcf
Number of TCA beds	3
Static sphere height per bed[*]	5 in.
Percent solids recirulated	15

Hold tank configurations tested were:

- Single hold tank at 12 minutes residence time.
- Three hold tanks in series at 12 minutes total residence time (4.3, 2.2, and 5.5 minutes).
- Three hold tanks in series at 9 minutes total residence time (3.8, 1.9, and 3.3 minutes).

Results of the utilization testing for these 3 configurations are plotted as stoichiometric ratio versus scrubber inlet liquor pH in Figures 5-8, 5-9, and 5-10, respectively. Sight drawn averages from Figures 5-8 through 5-10 are compared in Figure 5-11. For the runs with 3 tanks in series, there was no significant difference between 9 minutes and 12 minutes residence time. However, there was a distinct difference

---

[*] Initially 6-gram hollow TPR (thermoplastic rubber) spheres were used. These were subsequently replaced with 6.5-gram nitrile-PVC solid foam spheres.

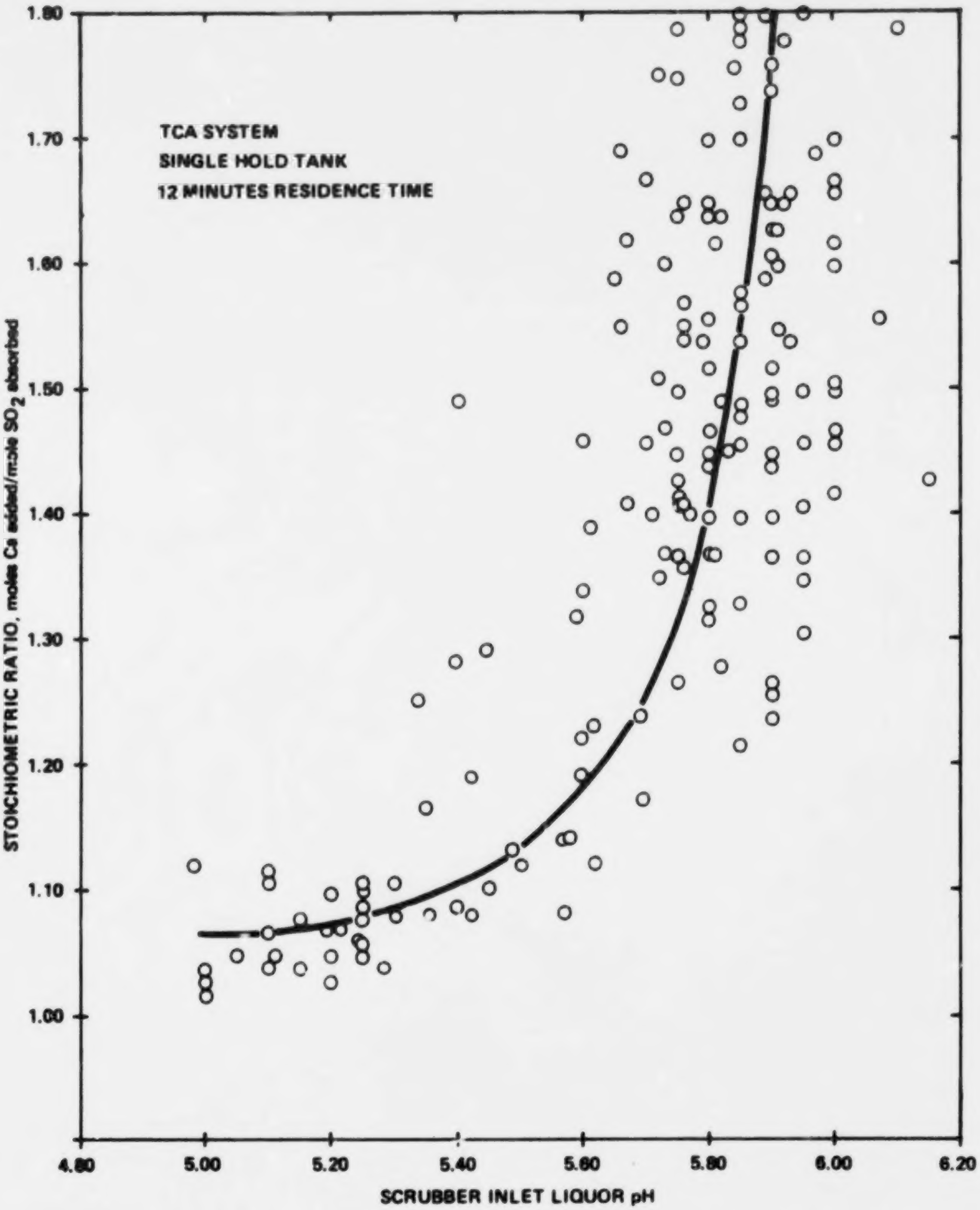


Figure 5-8. Stoichiometric Ratio versus Scrubber Inlet Liquor pH for a Single Hold Tank at 12 minutes Residence Time

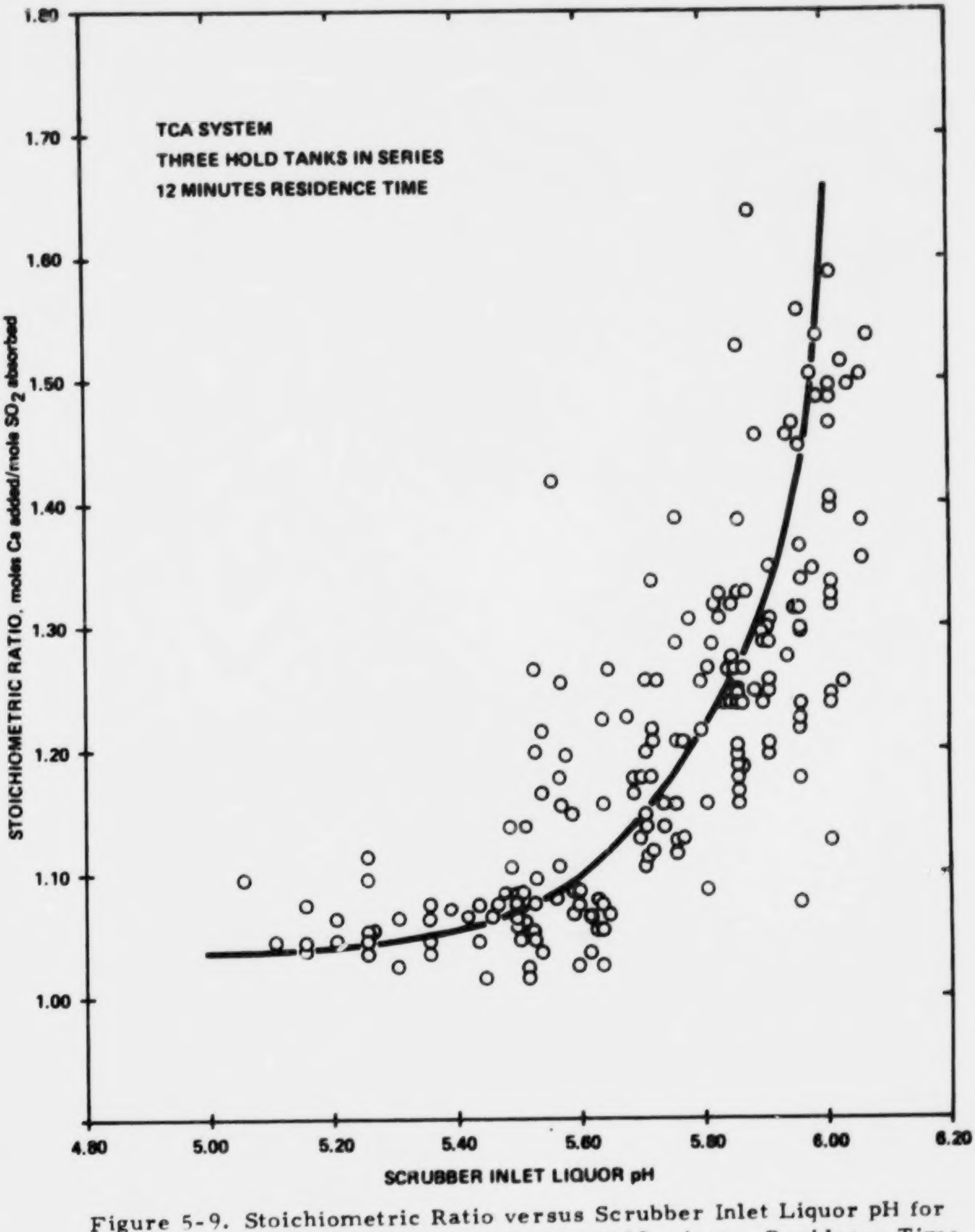


Figure 5-9. Stoichiometric Ratio versus Scrubber Inlet Liquor pH for Three Hold Tanks in Series at 12 minutes Residence Time

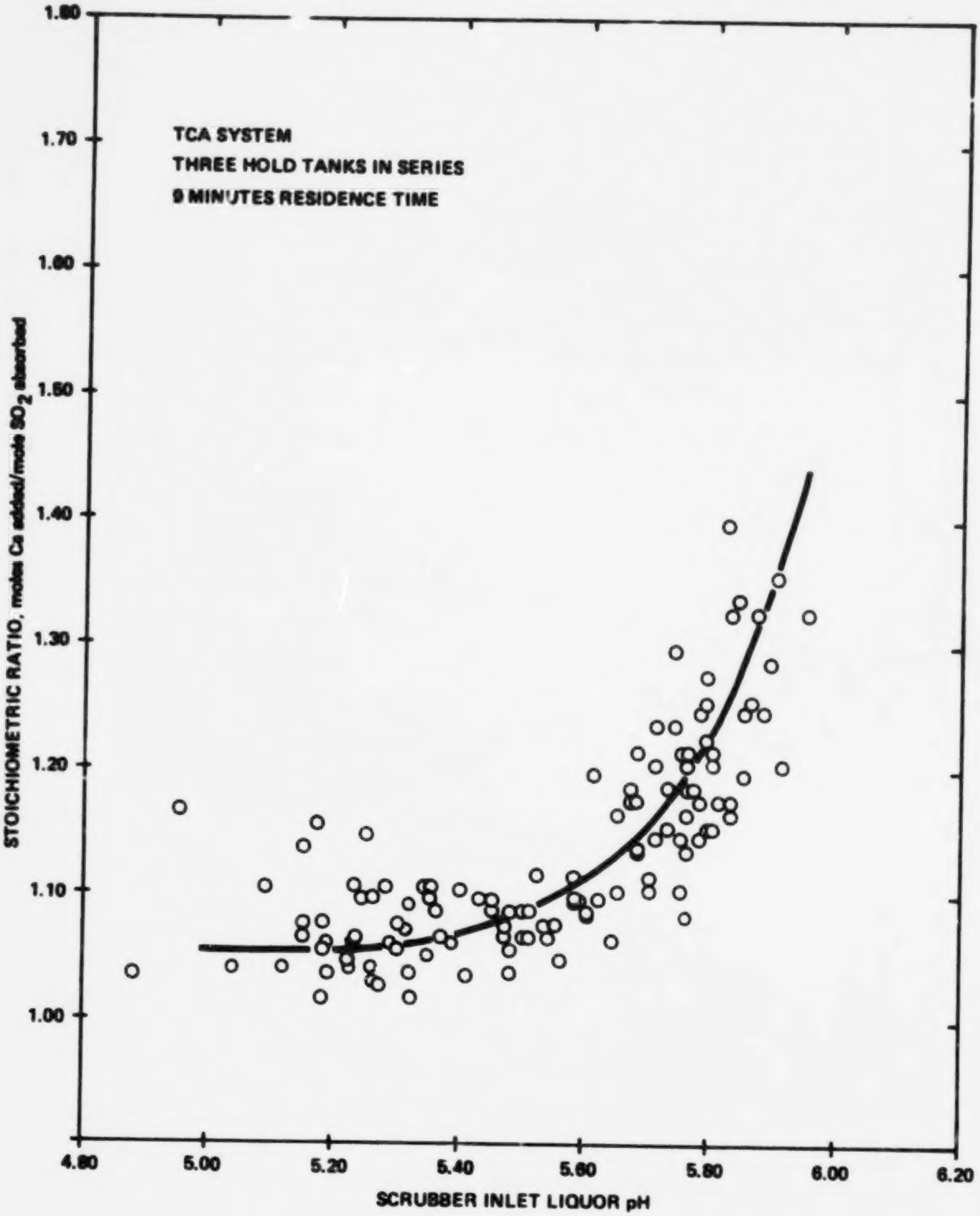


Figure 5-10. Stoichiometric Ratio versus Scrubber Inlet Liquor pH for Three Hold Tanks in Series at 9 minutes Residence Time

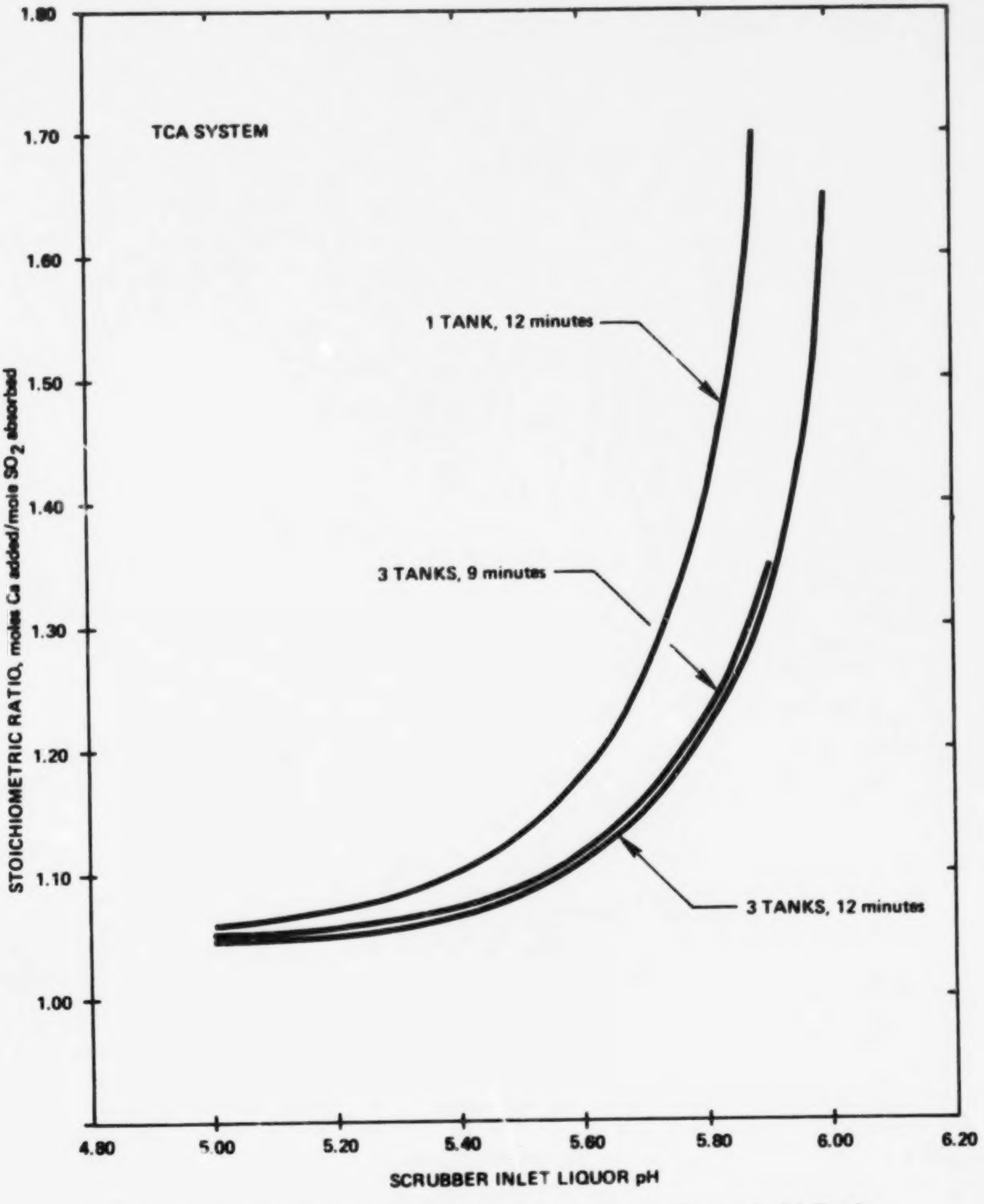


Figure 5-11. The Effect of Scrubber Inlet Liquor pH and Hold Tank Configuration on Stoichiometric Ratio

between operation with a single tank and with 3 tanks in series at pH levels greater than about 5.0. For example, at a scrubber inlet liquor pH of 5.6, stoichiometric ratio averaged 1.19 with a single hold tank as compared with 1.11 with 3 tanks in series, a 6 percent improvement in limestone utilization. At higher pH, the improvement was greater. For instance at 5.8 pH the improvement in utilization was 14 percent.

The improvement in utilization with 3 tanks in series can also be seen in Figure 5-12 where SO_2 removal is plotted versus stoichiometric ratio for an inlet gas SO_2 concentration range of 2500 to 3500 ppm. For example, at 85 percent SO_2 removal and 12 minutes total residence time, the stoichiometric ratio with 3 tanks in series averaged about 15 percent lower than with a single tank.

Data for 9 minutes residence time with 3 tanks in series were not included in Figure 5-12 because the TPR spheres in the TCA were replaced with nitrile-PVC solid foam spheres during the 9 minute testing. Unfortunately, the effect of 9 minutes residence time on SO_2 removal was confounded with the effect of the new bed of spheres on SO_2 removal.

5.3 CONCLUSIONS

Limestone utilization in the venturi/spray tower and TCA systems normally varied from about 60 percent at a scrubber inlet liquor pH of 6.0 to about 95 percent at a scrubber inlet pH of 5.2. Operation at reduced scrubber liquor pH, however, inherently causes a reduction in SO_2 removal efficiency.

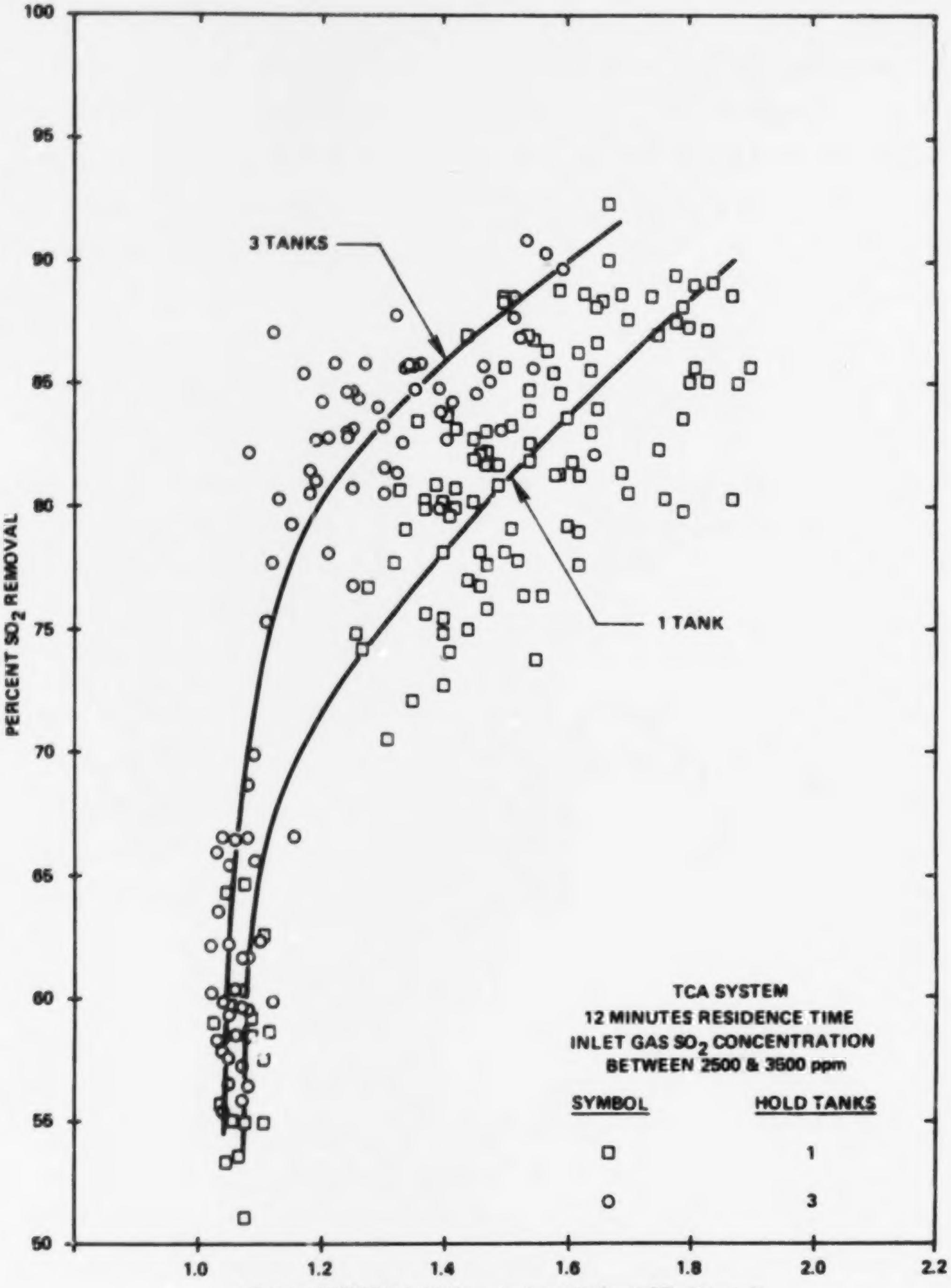


Figure 5-12. The Effect of Stoichiometric Ratio and Hold Tank Configuration on Percent SO_2 Removal

For the venturi/spray tower system with a single effluent hold tank, limestone utilization was not affected by a change in residence time from 20 to 12 minutes over the measured range of scrubber inlet liquor pH. Limestone utilization declined, however, at 6 minutes residence time for scrubber inlet liquor pH's greater than about 5.6.

For the TCA system at 12 minutes total effluent residence time and at scrubber inlet liquor pH's greater than about 5.0, higher limestone utilization is achieved with 3 hold tanks in series than with a single hold tank.

Section 6

EFFECT OF ALKALI UTILIZATION ON MIST ELIMINATOR OPERABILITY

Data showing the effect of alkali utilization on mist eliminator operability are presented in this section.

6.1 SUMMARY OF MIST ELIMINATOR OPERABILITY DURING LIME AND LIMESTONE RELIABILITY TESTING

During lime testing on the venturi/spray tower system, reliable operation was achieved for the 3-pass, open-vane, 316 stainless steel chevron mist eliminator (see Section 3). These tests were conducted at spray tower gas velocities up to 9.4 ft/sec (maximum attainable) and recirculated slurry solids contents up to 15 percent. The lime utilization for these tests was about 90 percent. Both the topside and the bottomside of the mist eliminator were washed intermittently with fresh water. Only about one-half of the makeup water available during closed liquor loop operation was required for mist eliminator wash.

During limestone testing on the TCA system with the wash tray and 6-pass, closed-vane, chevron mist eliminator in series (see Section 4), reliable operation of the mist elimination system was achieved only at 8.6 ft/sec scrubber gas velocity (5.6 ft/sec superficial gas velocity in the wash tray and mist eliminator areas). The limestone utilization ranged from 60 to 75 percent.

In view of the success of the spray tower mist eliminator operation with lime, a new mist eliminator of similar design (single stage, 316 stainless steel, 3-pass, open-vane, chevron type) was installed in the TCA in July 1975 for testing with limestone slurry. However, poor operability with this mist eliminator was experienced with intermittent topside and bottomside wash with raw water. For example, at a scrubber gas velocity of 9.4 ft/sec (6.1 ft/sec superficial gas velocity in the mist eliminator area), the chevron mist eliminator became about 50 percent restricted with soft solids after only 60 hours of operation (Run 554-2A). The limestone utilization was about 63 percent at the scrubber inlet liquor pH of about 6.0. This failure in the TCA, as compared with the successful operation in the spray tower with lime using a similar mist eliminator design and washing scheme, was initially attributed to the differences in the physical design and the pattern of mist generation within the two scrubbers.

During subsequent testing of the TCA system at scrubber inlet pH from 5.7 to 6.0 (limestone utilizations from 58 to 74 percent), stable conditions with less than 10 percent solids restricted were achieved for the mist elimination system by continuous washing of the underside of the chevron with diluted clarified liquor. For these low utilizations, only the mist eliminator blades continuously contacted by wash water could be maintained free of solids restriction. It was necessary, therefore, to eliminate shadowing by mist eliminator supports and to avoid maldistribution of gas caused by observation ports and corrosion coupon racks. In Run 559-2A, the chevron mist eliminator was only 7 percent restricted with soft solids after 384 operating hours using continuous underwash with diluted clarfied liquor at 0.4 gpm/ft^2 and intermittent topside wash with raw water.

6.2 MIST ELIMINATOR OPERABILITY DURING LIMESTONE UTILIZATION TESTING

In October 1975, limestone utilization testing was started on both the TCA and venturi/spray tower systems (see Section 5), concurrent with the on-going limestone mist eliminator testing in the TCA. Tables 6-1 and 6-2 summarize the results of mist eliminator operability during these tests through January 1976. These tables list the average scrubber inlet pH, average limestone utilization and stoichiometric ratio, mist eliminator wash scheme, and percent of the mist eliminator passage restricted by solids deposit at the end of each run. All of the available makeup water was used for tests with continuous mist eliminator bottom wash, while only about one-half of the available makeup water was used for tests with intermittent bottom wash.

During venturi/spray tower limestone Runs 701-1A and 702-1A (see Table 6-1) with intermittent topside and bottomside raw water wash, the mist eliminator was heavily restricted by soft solids within 2 to 3 days. This was unexpected since earlier operation with lime slurry under identical operating conditions was successful. The limestone utilization for these two runs was only 69 percent, as compared to 90 percent normally obtained with lime operation. Subsequently, during Run 703-1A, the limestone utilization was increased to 93 percent by dropping the average scrubber inlet pH to 5.2. The mist eliminator was found to be essentially clean (1 percent restricted) after 319 operating hours with intermittent underside and topside raw water wash. The average SO_2 removal for Run 703-1A was only 58 percent, as compared with 87 percent for the previous tests.

Table 6-1

SUMMARY OF VENTURI/SPRAY TOWER LIMESTONE UTILIZATION AND MIST ELIMINATOR TESTS

All runs made with a 316 stainless steel, 3-pass, open-vane, chevron mist eliminator. Run conditions: 15 wt % solids in recirculated slurry, 9.4 ft/sec spray tower superficial gas velocity, 21 gal/mcf liquid-to-gas ratio in venturi, 50-57 gal/mcf liquid-to-gas ratio in spray tower.

Run No.	Single Tank Res. Time, min	Avg. Scrubber Inlet pH	Avg. Stoich. Ratio	Avg. Percent Limestone Utilization	Avg. Percent SO_2 Removal[g]	Mist Eliminator Wash Scheme		Run Hours	Hours Since Cleaning Mist Eliminator	Percent Mist Eliminator Restriction
						Top	Bottom			
701-1A	20	5.9	1.45	69	88	Intermittent[a]	Intermittent[b]	73	73	50-60
702-1A	20	5.8	1.45	69	87	↓	↓	60	60	45-50
703-1A	20	5.2	1.07	93	58	↓	↓	319	319	1
704-1A	20	5.8	1.45	69	86	↓	↓	66	385	45-50
705-1A	20	5.7	1.25	80	84	↓	↓	136	136	17-20
706-1A	12	5.2	1.06	94	58	↓	↓	180	180	1
707-1A	12	5.7	1.30	77	83	Intermittent[c]	Intermittent[d]	118	298	10-15
708-1A	12	5.6	1.20	83	77	↓	Intermittent[d]	138	436	15-20
709-1A	12	5.9	1.35	74	83	↓	Continuous[e]	134	134	<1
710-1A	12	6.0	1.50	67	91	↓	↓	234	368	5-7
711-1A	6	5.7	1.30	77	81	↓	↓	144	512	-
711-1B	6	5.6	1.40	71	82	↓	↓	71	583	5-7
712-1A	6	5.8	1.50	67	87	↓	↓	119	702	10-15
712-1B	6	Depletion[f]	-	-	-	↓	↓	18	720	-
713-1A	6	5.2	1.10	91	69	↓	↓	52	772	10

(a) Intermittent, sequential top wash with makeup water at 0.53 gpm/sqft for 4 min/8 hr/section.
(b) Intermittent, full face bottom wash with makeup water at 1.5 gpm/sqft for 6 min/4 hr.
(c) Intermittent, sequential top wash with makeup water at 0.53 gpm/sqft for 3 min/hr/section.
(d) Intermittent, full face bottom wash with makeup water at 1.5 gpm/sqft for 4 min/hr.
(e) Continuous, full face bottom wash with diluted clarified liquor at 0.4 gpm/sqft.
(f) A limestone depletion run is conducted without limestone addition during SO_2 absorption. The scrubber inlet liquor pH is allowed to drop from about 5.9 to 4.8.
(g) SO_2 removals are for 2500 to 3500 ppm inlet gas SO_2 concentration.

Table 6-2

SUMMARY OF TCA LIMESTONE UTILIZATION AND MIST ELIMINATOR TESTS

All runs made with a 316 stainless steel, 3-pass, open-vane, chevron mist eliminator. Run conditions: 15 wt % solids in recirculated slurry, 12.5 ft/sec TCA superficial gas velocity, 50 gal/mcf liquid-to-gas ratio. TCA configuration: 3 beds with 5 inches per bed of 6-gram TPR spheres for Runs 562-2A through 569-2A and 6.5-gram nitrile foam spheres for Runs 569-2B through 582-2A.

Run No.	No. Tanks	Total Res. Time, min	Avg. Scrubber Inlet pH	Avg. Stoich. Ratio	Avg. Percent Limestone Utilization	Avg. Percent SO_2 Removal[e]	Mist Eliminator Wash Scheme		Run Hours	Hours Since Cleaning Mist Eliminator	Percent Mist Eliminator Restriction
							Top	Bottom			
562-2A	1	12	5.9	1.6	63	83	Intermittent[a]	Continuous[b]	495	155	2
							↓	↓		332	5-7
							↓	↓		495	7-9
562-2B	1	12	5.7	1.4	71	81	↓	↓	134	629	7
563-2A	1	12	5.9	1.7	59	86	↓	↓	182	811	7
564-2A	1	12	5.2	1.06	94	58	↓	↓	113	924	3
565-2A	3	12	5.25	1.05	95	58	↓	↓	109	109	0
566-2A	3	12	5.9	1.25	80	83	↓	↓	166	275	0
567-2A	3	12	6.0	1.4	71	84	↓	↓	138	413	2
568-2A	3	12	5.5	1.06	94	63	↓	Intermittent[c]	162	575	1
569-2A	3	9	5.5	1.08	93	69	↓	↓	66	641	<1
569-2B	3	9	5.5	1.08	93	64	↓	↓	97	738	<1
570-2A	3	9	5.75	1.18	85	72	↓	↓	144	882	<1
571-2A	3	9	5.8	1.25	80	73	↓	↓	96	978	<1
571-2B	3	9	Depletion[d]	-	-	-	↓	↓	11	989	-
572-2A	3	9	5.25	1.09	92	57	↓	↓	154	1143	-
573-2A	3	12	5.5	1.08	93	66	No wash	↓	45	1188	<1
573-2B	3	12	Depletion	-	-	-	↓	↓	28	1216	-
574-2A	3	9	Depletion	-	-	-	↓	↓	12	1228	40
575-2A	3	12	5.55	1.15	87	69	↓	↓	47	47	-
576-2A	3	12	5.7	1.17	85	73	↓	↓	112	159	-
576-2B	3	12	Depletion	-	-	-	↓	↓	3	162	3
577-2A	3	12	5.8	1.25	80	83	Intermittent[a]	↓	159	159	20
578-2A	3	12	Depletion	-	-	-	↓	↓	9	168	-
579-2A	3	9	5.25	1.05	95	61	↓	↓	62	230	-
580-2A	3	9	Depletion	-	-	-	↓	↓	12	242	10
581-2A	1	12	5.5	1.15	87	80	↓	↓	164	406	-
582-2A	1	12	Depletion	-	-	-	↓	↓	18	424	2

(a) Intermittent, sequential top wash with makeup water at 0.55-0.83 gpm/sq ft for 3 min/hr/section.

(b) Continuous, full face bottom wash with duluted clarified liquor at 0.4 gpm/sq ft.

(c) Intermittent, full face bottom wash with makeup water at 1.5 gpm/sq ft for 4 min/hr.

(d) A limestone depletion run is conducted without limestone addition during SO_2 absorption. The scrubber inlet liquor pH is allowed to drop from about 5.9 to 4.8.

(e) SO_2 removals are for 2500 to 3500 ppm inlet gas SO_2 concentration.

Subsequent venturi/spray tower Runs 704-1A through 708-1A confirmed the observation that reliable mist elimination operation could only be obtained at high alkali utilization (greater than about 90 percent), with intermittent underside and topside wash using raw water. Runs 709-1A through 711-1B also showed that for utilization less than about 80 percent, continuous bottomside wash with diluted clarified liquor could reduce or limit the amount of soft solids deposition on the mist eliminator vanes.

Testing with the TCA system similarly confirmed the strong effect of limestone utilization on mist eliminator reliability. Runs 562-2A, 562-2B and 563-2A (see Table 6-2) were conducted at average scrubber inlet pH's of 5.7 to 5.9 with limestone utilizations from 59 to 71 percent. The mist eliminator was washed intermittently with fresh water on the topside, and continuously with diluted clarified liquor on the bottomside. The mist eliminator restriction increased to 7-9 percent during the first 500 hours of operation and appeared to level out at 7 percent restriction after 811 hours at the end of Run 563-2A. Following these tests, the scrubber inlet pH was dropped to 5.2 (Run 564-2A) and the operation continued for additional 113 hours. At the end of Run 564-2A the mist eliminator restriction decreased to 3 percent from the 7 percent at the start of the run. The limestone utilization during Run 564-2A averaged 94 percent.

As was discussed in Section 5, operation at reduced scrubber liquor pH and high utilization inherently causes a reduction in SO_2 removal efficiency. This reduction in efficiency can be compensated by: (1) increasing the slurry flow to the absorber, (2) increasing the TCA packing height and gas phase pressure drop, or (3) adding MgO to the scrubber slurry. Another way of increasing SO_2 removal efficiency while maintaining high limestone utilization is to operate with three hold tanks in series. TVA is presently conducting an economic evaluation of the proposed schemes for improving SO_2 removal efficiency while maintaining high limestone utilization.

TCA Runs 565-2A through 580-2A were operated with three hold tanks in series (see Table 6-2). As with the venturi/spray tower system, continuous bottomside wash with diluted clarified liquor and intermittent topside wash with raw water limited the amount of soft solids deposition at low utilization (Run 567-2A). Also, the mist eliminator was kept completely free of solids with continuous bottomside wash at high utilization (Runs 565-2A and 566-2A). As expected, with intermittent bottomside and topside wash at high alkali utilization (Runs 568-2A through 573-2A), the chevron mist eliminator remained essentially free of solids. After a total of 1188 operating hours with continuous and intermittent bottomside wash and intermittent topside wash, the chevron mist eliminator in the TCA was less than one percent restricted with soft solids (see Run 573-2A in Table 6-2).

A further example of a mist eliminator which has become less restricted during operation at high utilization can be seen in Runs 577-2A through 582-2A. At the conclusion of Run 577-2A the mist eliminator was 20 percent restricted by soft solids and after an additional 226 hours of operation at 95 and 87 percent utilization (Runs 579-2A and 581-2A), the mist eliminator restriction decreased to 2 percent.

6.3 CONCLUSIONS

The following general conclusions can be drawn for the performances of the single stage, 316 stainless steel, 3-pass, open-vane, chevron mist eliminator in both the TCA and venturi/spray tower systems with lime and limestone:

- The reliability of the mist elimination system is a strong function of alkali utilization.
- For high alkali utilization (greater than about 85 percent), the mist eliminator can be kept free of solids deposits with intermittent fresh water top wash combined with either intermittent fresh water bottom wash or continuous bottom wash with diluted clarified liquor. Intermittent top and bottomside fresh water wash may be required in closed liquor loop operation due to restrictions in the allowable raw water makeup to the scrubber system.
- For alkali utilization less than about 85 percent, intermittent top and bottomside wash with fresh water does not limit solids accumulation. However, for these conditions, a continuous bottom wash with diluted clarified liquor used in combination with an intermittent topside fresh water wash can limit soft solids buildup to less than 10 percent restriction within the mist eliminator.
- Operation for a period of time at high alkali utilization (greater than about 90 percent) may, in certain instances, clean up an already fouled mist eliminator.

The effect of high alkali utilization on decreasing soft solids retention within a mist eliminator should have direct application to commercial scrubber design and operation. It may be possible to change operating conditions in existing scrubbers to increase limestone utilization. High utilization can be designed into new scrubber systems.

Section 7

REFERENCES

1. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Summary of Testing through October 1974, EPA Report 650/2-75-047, June 1975.

2. Bechtel Corporation, EPA Alkali Scrubbing Test Facility: Advanced Program-First Progress Report, EPA Report 600/2-75-050, September 1975.

3. R. H. Borgwardt, "Increasing Limestone Utilization in FGD Scrubbers," presented at the Sixty-Eighth Annual Meeting of the AIChE, Los Angeles, November 16-20, 1975.

4. R. H. Borgwardt, "IERL-RTP Scrubber Studies Related to Forced Oxidation of Sludge," presented at the IERL Symposium on Flue Gas Desulfurization, New Orleans, March 8-11, 1976.

DUQUESNE LIGHT COMPANY
ELRAMA AND PHILLIPS POWER STATIONS
LIME SCRUBBING FACILITIES

R. Gordon Knight, Superintendent, Technical Services
Steve L. Pernick, Jr., Manager, Environmental Affairs

Duquesne Light Company
Pittsburgh, Pennsylvania

ABSTRACT

The Elrama and Phillips Power Stations are coal burning facilities having net generating capabilities of 494 Mw and 387 Mw, respectively. Both plants are located in the southwest portion of Pennsylvania within a radius of approximately 15 miles of the city of Pittsburgh.

Venturi scrubbing systems using lime were installed at both facilities for dust removal, with one of the scrubbers at the Phillips facility equipped with a dual stage venturi scrubber to be used and tested as the prototype for sulfur dioxide removal. Startup of a portion of the Phillips scrubber system began July, 1973. This system is designed to scrub the headered gases from all six boilers of the plant, and discharge them to a new 340 ft. stack after reheating them from 120^o to approximately 140^o with direct fired No. 2 oil. The scrubber system consists of four trains. The first train is the prototype for sulfur dioxide removal and consists of two identical venturi scrubbers in series. Each of the remaining three trains consist of a single venturi scrubber. A series of tests were conducted at this facility to determine the most feasible means of complying with the SO_2 emission limitations.

Early operation of the Phillips scrubber system revealed several problems, such as equipment erosion and corrosion, acid condensate leakage in the new stack, and stress problems in the ID fans, which caused outages of the scrubber system. After an extended outage, the scrubber system resumed partial operation in March, 1974, and has been in continuous operation since that time with a varying number of boilers connected to the scrubber system and with a varying number of scrubber trains in service. For approximately a five month period in 1975, all six boilers were connected to the scrubber system. During this period, the scrubber availability was approximately 67%.

Our operating experiences have revealed many problems, such as scaling and deposits in the scrubber, ID fan deposits and high stresses,

a crack in an ID fan shroud, recycle pump erosion/corrosion, high water inventory, sludge disposal problems, and lime addition problems.

In August, 1975, the scrubber problems were adversely affecting the plant generating capability to the point where it became necessary to remove the largest boiler, No. 6 Boiler, from the scrubber system and return it to its original gas path. Operation has continued with this arrangement until the present time.

Because of the serious nature of the problems with the Phillips scrubber system, startup of the Elrama scrubber system was delayed to permit study and resolution of the problems at Phillips. The Elrama scrubber system, consisting of five single stage scrubber trains serving four boilers, was placed in partial operation with one boiler connected to it on October 26, 1975. A second boiler was connected on February 4, 1976, and operation has continued in that manner to the present time.

Tests conducted on the Phillips scrubber system indicate SO_2 removal efficiencies using high calcium lime of approximately 90% with dual stage scrubbing and 50% with single stage scrubbing. Tests using a magnesium modified lime (magnesium oxide of 8-10%) indicate removal efficiencies of approximately 83% with single stage scrubbing. The results of the magnesium modified lime tests indicate an ability to comply with the SO_2 emission limitations with single stage scrubbing.

More than $66 million has been spent to date on both scrubber systems with an additional $20 million to be spent within the next several years to complete the installations for removing sufficient SO_2 to comply with the regulations. Operating costs of the Phillips Station have increased approximately 35% due to the operation of these scrubbers.

Although many problems have been and are being resolved, others still remain, and consequently, an assessment of the entire system operation, reliability, and practicability is inconclusive at this time.

DUQUESNE LIGHT COMPANY
ELRAMA AND PHILLIPS POWER STATIONS
LIME SCRUBBING FACILITIES

INTRODUCTION

Duquesne Light Company, an investor-owned electric utility, serves about one-half million customers in Southwestern Pennsylvania and has a net generating capability of approximately 2500 Mw. This capability is generated by combustion turbines in simple and combined cycle modes, by nuclear plants, and by coal fired power stations. The Company is sole owner and operator of three coal fired stations, two of which have been retrofitted with wet scrubbers using lime as a reagent.

The history, description, and early operating experiences of the scrubbers at our Phillips Power Station were described in a paper at the EPA Flue Gas Desulfurization Symposium in November, 1974. A second paper presented to APCA in June, 1975 described the Phillips operating experiences during the period March, 1974 through April, 1975. This paper will update the Phillips experiences, giving results of tests that have been and are being conducted, and it will describe the startup and operating experiences of the Elrama Power Station scrubbing system for the period extending through January, 1976.

DESCRIPTION OF SYSTEMS

Phillips Power Station

Four trains of wet venturi-type scrubbers have been installed at the 387 Mw Phillips Power Station at a cost of approximately $103 per Kw. Gibbs & Hill, Inc. was engaged as architect-engineer for the entire project, with battery limits of the Chemico Corporation confined to the scrubbers and associated pumps and controls between the inlet hot gas duct manifold and the exit wet gas header, including the reheater but excluding the new induced draft fans.

The four trains are located downstream of existing in series mechanical collectors and electrical precipitators installed on each of the six pulverized coal fired boilers. Three of the trains are single stage venturi scrubbers originally intended for particulate removal. The fourth train is dual stage and was the prototype for determining the feasibility of two stage scrubbing for compliance with SO_2 emission limits. A test program on the prototype and the single stage scrubbers was used to determine what additional equipment or process will be necessary to enable us to meet regulatory requirements for SO_2 removal, which are essentially 83% SO_2 removal efficiency using 2% sulfur coal.

Each train is equipped with a new wet-type ID fan. A new common duct directs emissions from all boilers to the scrubber plant where they can be sent to any or all of the trains. The scrubbed gas is exhausted through a common duct and an oil fired reheater to a new ground supported stack which consists of a concrete envelope, a 30 inch annulus, and an inner acid-resistant brick stack.

Slaked quicklime is added to the lower cone of each of the scrubber vessels to neutralize the recirculating liquor, which, in single stage scrubbing with high calcium lime, reacts with about 50% of the SO_2 in the flue gas. A liquor bleed flow of about 4% is sent to one or both of two thickeners for solids removal. The overflow is returned to the system, and the underflow is piped to the sludge treatment system.

Startup of a portion of the Phillips scrubber system began July, 1973. Several problems then developed causing outages of the scrubber system, and after an extended outage, the scrubber system was returned to service in March, 1974. The system has been operating continuously since that time with a varying number of boilers connected to the scrubber system, and with a varying number of scrubber trains in service.

Elrama Power Station

The scrubber facility at the 494 Mw Elrama Station is identical to that at Phillips in most ways. Gibbs & Hill is the architect-engineer, Chemico scrubbers have been installed, and the battery limits are the same. Mechanical and electrical dust removal equipment remove most particulates from the boiler emissions, and the gas to and from the scrubbers is headered in the same way. Unlike Phillips, there are four boilers, each with its own turbine generator. Five single stage scrubbers were installed, and it was intended that the knowledge gained from the test program at Phillips would be applied to Elrama to enable compliance with the regulations.

The first Elrama scrubber was not placed in service until October 26, 1975. It had been scheduled for an earlier startup date; however, the severity and number of problems encountered at Phillips made it prudent to delay startup until many of the problems at Phillips were resolved and the necessary modifications made at both stations. The decision to delay the Elrama startup has proved to be a wise one.

PHILLIPS OPERATING EXPERIENCE

A number of problem areas were described in our previous papers. Some of these have been alleviated, and others still persist. The present status of a number of these is described below.

ID Fans

Encouraging progress has been made in this area. The second repair of the original defective welds with Inconel 112 has proved successful and resulted in very few minor or no weld repairs at each 1500 hour inspection. Relocation of spray nozzles with the use of a new type (Bete fog nozzle No. TF16FC) has practically eliminated the deposit accumulations, which caused fan imbalance and under which pitting attack occurred.

A problem still exists with pitting of the 316 SS sleeves on the fan shafts. Recent trial use of a coating of Neoprene-29 appears encouraging. The rubber lining on the housings needs periodic minor repairs.

During a routine inspection of a fan in the last week of January, a crack was found in one of the shrouds. The crack extended about 4 inches from the outer periphery and was about 1/16 of an inch wide at that point. Both shrouds will be checked by non-destructive methods. The defective area will be removed for examination. Repairs will be made, and the fan returned to service. This fan has experienced about 11,000 service hours. The other fans will be inspected and checked by non-destructive methods.

Recycle Pumps

The corrosion/erosion problem with these pumps continues. As indicated in our previous paper, we are following a trial program involving a number of different impeller and wear ring materials. Although we have still not obtained the desired life from any of the materials we have tried, there are indications that a CD4MCu impeller with a 26% chrome-iron wear ring is the best combination. Service life with these two materials has been up to 4700 hours, compared to the 3000 hours with the original Carpenter 20. Two rubber lined pumps are on hand for trial installation at the Elrama Station.

Bleed Valves

Previous problems with rubber lined plug-type bleed valves have apparently been solved by the substitution of pinch-type valves.

Deposits

Experience has shown that complete cleaning of a scrubber vessel is required about every 1400 service hours. Each cleaning with minor maintenance requires 1400-1700 man hours. Approximately 20-80 tons of deposits are removed during each cleaning. These deposits result even though the pH of the recycle liquor is maintained in the 6-7 range for about 95% of the time. We do not yet have redundant lime feed or automatic pH control. These are to be forthcoming.

The main troublespot for deposits is in and around the throat dampers. Despite programmed, automatic exercising of the dampers, deposits accumulate to the extent that the throat pressure drop increases from a normal 6 inches to 12 inches or more after about 1400 hours of service. This high delta p decreases the amount of gas that can be scrubbed to such an extent that generating capability of the power plant is reduced and a cleaning outage is necessary. The scrubber manufacturer had installed and recommended that we keep open a 1-1/2 inch valved vent on each side of the damper housings in an effort to sweep the areas clear of deposit with outside air. This had no effect. To obtain more air sweep, we next removed eleven 3/4 inch plugs from openings on the sides and tops of four of the twelve damper housings on one of the vessels. This, too, had no effect. Earlier we had introduced a continuous flow of water through the same 3/4 inch openings with little or no benefit.

Deposits accumulate in and around the twelve tangential nozzles in each vessel. This results in an uneven flow of scrubbing liquor around the upper cone. Chemico recommended that alternate nozzles in one of the vessels be blanked off to see if the increased velocity through the others would prevent the accumulations. After 1506 service hours, this appears to be partially effective in the vessel where it was tried.

Sizable deposits form in the lower cone, where the lime discharges into the vessel through an open end pipe. Chemico recommended the use of an angled nozzle at this location. This is currently being implemented.

Appreciable deposits are also found on the lip under the upper cone. Chemico is conducting scale model studies in an effort to determine the cause and possible corrective measures.

Comments on all these deposit problems should be considered in conjunction with the results obtained in our magnesium modified lime tests. These will be discussed later.

PHILLIPS SLUDGE TREATMENT AND DISPOSAL

Previous reports have discussed our first-of-a-kind experience with Dravo's Calcilox as a fixating agent. Early unsuccessful use resulted from our inability to maintain one or a combination of (1) the required 10% addition (by dry weight of sludge), (2) warm ambient temperature, (3) necessary curing time, and (4) pH of 10-11. None of these would be critical if we could add the fixative and then let the mixture remain in place. However, we have only three "curing" ponds, each of about 6000 yd^3 capacity, and we fill each one in 10-14 days. Furthermore, to be prepared for unexpected emergencies, such as the draining of a scrubber or a thickener, we must have an empty or near empty pond at all times. Consequently, we must excavate ponds within about 4 days of filling which precludes the necessary curing time. The problems involved with hauling "soup" in open dump trucks are obvious. When requirements are satisfied, we obtain excellent fixation with Calcilox. The original system introduced the additive in slurry form. In April, 1975, we installed a dry additive system, fabricated by and rented from Dravo. This was intended to be the primary means of treatment with redundancy provided by the original wet system. For a short period, we experienced somewhat more constant additive feed, but frequently both systems were out of service. The dry system is a modified bulk concrete plant and leaves much to be desired in feeding Calcilox. Because of excessive maintenance requirements of the dry system, we are now using only the wet system. Recent experiences with lower than normal sludge pH will be described later.

In June, 1975, IU Conversion Systems (IUCS) installed a pilot plant at Phillips to process sludge by their POZ-O-TEC system. The feed to their system was a portion of our thickener underflow containing about 30-40% solids, consisting of about equal portions of fly ash and $CaSO_3/CaSO_4$. The processed product contained about 70% solids and could be transported and compacted immediately. About 400 tons of processed material were produced and placed in an isolated area in our land fill operation. Additional experience with this process is detailed later.

Disposal tests on landfill areas are also being made by our ash hauling contractor, U. S. Utilities Service. They are mixing scrubber sludge with various quantities of fly ash and bottom ash, then spreading and compacting. Observations of stability and leachability are being made.

Incidentally, the three companies mentioned in the preceding are the active bidders on our long term waste management contract, now under negotiation. We are presently reviewing their proposals for long term sludge management in hopes of improving our disposal techniques and in preparation for the larger quantities and different type of sludge which will be produced when SO_2 compliance is achieved.

Our leachate tests have been previously described. Those which involved the membrane lined disposal areas are continuing, but freezing temperatures prevent sampling at this time of year. Those involving various mixtures in lightweight plastic pools have been discontinued because of failure of the pools. We are about to start another series using heavy plastic "tote" boxes. Our most recent test results show that the fixed sludges are continuing to leach higher concentrations of total dissolved salts than reported by others.

PHILLIPS SCRUBBER SYSTEM AVAILABILITY

Reports by others on availability of the Phillips scrubbers have been hopelessly confused and unduly optimistic because of the header design of the plant. Until all boilers were connected to the scrubbers, there was at least one spare train and as much as 100% spare scrubber capacity. Therefore, having a train out of service for maintenance did not reduce the capability of the scrubber plant. Until all six boilers were connected, meaningful availability factors were not available. On March 17, 1975, the last boiler was connected, and all four scrubber trains were required to be in service. Operation in that mode continued until August 4, 1975 when it was necessary to remove the No. 6 Boiler from the scrubber system because the pH in the system could not be maintained, and deposits were becoming unmanageable to the point where scrubber outages were reducing the station generating capability. In the four and a half month period of total scrubbing, availability averaged 67%.

PHILLIPS SCRUBBER SYSTEM TESTS

In addition to our continuing performance and sludge tests, several other tests are noteworthy.

Lamella Thickener

Because of the space limitations at both stations and because of the success of the phosphate industry in using the Lamella plate-type thickener, we obtained and operated a 100 gpm model at Phillips for four weeks. A polymer was added to the inlet stream from the scrubber bleed. The influent contained 2.5% to 3.0% solids, and the Lamella increased the concentration to the range of 17-20%. We require a 30-40% concentration. The overflow quality was excellent - less than 150 ppm. The manufacturer said that use of a larger bottom hopper would have given a greater concentration of solids in the underflow. No good evaluation of long term deposit accumulation and ease of removal can be given. The Lamella plates were disassembled after 17 days of continuous operation, and soft deposits of 1/4 to 1 inch in thickness were found near the bottom of the plates. These were easily removed from the plates, after removal from the housing, by medium pressure jets. Removal of deposit from in-place plates cannot be evaluated.

Additives

We have been plagued with black plumes from the scrubber stack at times of boiler upsets. Although we were originally told that the scrubbers would remove unburned carbon, this has not been the case. In an effort to clean up the plume, we tried four different wetting agents, all from the same supplier. None had any effect.

Magnesium Modified Lime

We ran a two and a half month test with magnesium modified lime in order to determine (1) if single stage scrubbing could remove sufficient SO_2 to enable compliance with regulations, (2) the effect on deposits, and (3) the change in sludge characteristics. Because of our limited lime capacity, it was necessary to reduce the quantity of emissions treated in the scrubbers. We did this by re-routing the emissions from a second boiler back to the original stack - this boiler was later taken out of service for major maintenance - and by limiting operation on a third to supplying emergency generation requirements. In order to slake the blended lime completely, a source of hot water was obtained so that slaking temperature was held in the 180-190°F range. Efforts were made to reduce the addition of fresh water in order to maintain system chemistry. An effective step was the use of thickener overflow return water to dilute the lime slurry after slaking. A not-so-effective step was the reduction of demister spray time. Deposits resulted until normal spray time was restored.

System requirements limited the duration of the test so that the optimum value of some operating parameters must be determined at a later time. However, the test did yield some valuable information.

It was determined that the required degree of SO_2 removal - approximately 83% - can be obtained with an MgO content of 8-10% in the lime with single stage scrubbing. This will eliminate the necessity to install any additional stages or scrubber vessels at both the Phillips and Elrama Stations. This will save approximately $10 million in capital costs.

The two and a half month test with modified lime resulted in 1612 service hours on one train and 1309 on another. After this length of service, the throat damper pressure drop in both trains was still 6 inches, the same as with newly cleaned dampers. With the normal high calcium lime, the pressure drop would have been at or approaching 12 inches, and an outage for cleaning would have been necessary. Inspection showed negligible deposit accumulation in all accessible locations, and the deposit on some previously scaled surfaces had been removed. For example, the maximum flow that could be obtained through the bleed line on one scrubber prior to the test was 420 gpm. This flow increased during the test to a value of 720 gpm at termination. On the basis of SO_2 removal efficiency and freedom from deposits, it is unfortunate that we cannot continue the use of the modified lime. As discussed later, this requires major modifications, about 18-24 months for construction and the expenditure of about $20 million to convert both the Phillips and Elrama scrubbers.

One of the less favorable effects of using magnesium modified lime was on the sludge, which did not settle as well in the thickeners. The solids concentration in the underflow decreased from 35-45% to 25-30%. Although a polymer was tried, time did not permit a complete evaluation of its effectiveness. The nature of the solids is important from two considerations: effect on sludge treatment schemes and on the size or number of additional thickeners needed for the next phase of construction. Steps were taken to supply sludge samples to the thickener builders and to the three bidders for our long term waste management contract. The results of the thickener tests have not yet been submitted to us. All three of the waste processors said that the sludge could be handled, with slight modification needed in some cases.

Another unexpected effect of our use of magnesium modified lime was a reduction in the pH of the sludge (8-9 instead of 10-11). This evidently resulted from better lime slaking, resulting in less unreacted lime in the sludge, and from higher utilization of lime by better SO_2 removal. The effect on sludge fixation was to leave us with soupy ponds to excavate and haul to the final disposal site. Dravo recommended the use of 15% (instead of 10%) Calcilox addition, and they added an "alkaline booster" to the Calcilox. We were not able to get the desired pH before the test ended, and, to make things worse, the increased SO_2 removal and consequent sludge filled the ponds faster and reduced the curing time. Dravo's representatives indicated that laboratory tests showed good fixation with the proper pH and curing time.

ELRAMA SCRUBBER STARTUP AND OPERATING EXPERIENCE

As previously noted, the first boiler was connected to the scrubbers on October 26, 1975. This boiler, No. 2, has an equivalent capacity of about 100 Mw, and the emissions can be handled by one scrubber. However, to insure reliability in case of a scrubber malfunction, two scrubbers are operated at partial load to protect the boiler and its associated turbine generator against a trip-off. We have had continuous operation of the boiler on the scrubber system through January, 1976 with the exception of two short outages. One was occasioned by inoperative throat dampers, originally thought to be a result of deposit buildup. Upon inspection, however, the cause was found to be damper scraper blades which had been installed with insufficient clearance. The other involved failure of the lime feeder belt in conjunction with a boiler outage. Four of the scrubber vessels have been in service in various combinations, and the service hours to January 31, 1976 are:

No. 1 - 1169 hours
No. 2 - 1508 hours
No. 3 - 976 hours
No. 4 - 838 hours

No. 5 has not been in service because it is being revised for the trial installation of rubber lined recycle pumps.

To this time, there have been no major problems. There have been many startup pains, not the least of which has been frozen pipes. Heat tracing has been and continues to be a problem at the Phillips Station, and it appears we have more of the same at Elrama. The most persistent problem, and one that could take the scrubber system and generating facilities out of service, has been experienced with the thickeners. This is a hardware and design problem associated with recirculation of the sludge within the thickeners to attain the 30-40% solids concentration. Fortunately, station personnel have recently found that our sludge treatment process (described below) can handle sludge with solids concentration as low as 15%; thus, recirculation is unnecessary. Each of the two thickeners has sufficient capacity for one boiler. Therefore, we have had spare capacity up to this point. When another boiler is connected to the scrubbers, a thickener outage will cause a boiler-turbine generator outage. Two boilers will be the total that can be placed on the scrubber system until additional construction is completed.

ELRAMA LIME SYSTEM

A report appearing in the literature in August, 1975 indicated that the Phillips scrubbers were the first in the United States to go in service using hydrated lime. That was a misstatement. However, it can now be said that the Elrama scrubbers were placed in service and are continuing with hydrate. However, we will change over to quicklime as soon as possible since it is more economical than hydrated lime.

The installed lime slaker, silo, and pumps had a feed capacity of 9 lb/minute with all scrubbers in service. Since we learned through experience at Phillips that this was inadequate for proper pH control, we did not want to go through a similar problem at Elrama, and because procurement of additional equipment would have caused an impossible delay, steps were taken to adapt the system to hydrated lime. The slaker was converted to a mixer, the silo feeder changed, and new transfer pumps installed. In addition, arrangements were made for the lime supplier to leave two loaded 20 ton trailers on site at all times. These, along with our 20 ton silo and a recently purchased 50 ton storage "blimp" and permanently installed lime blower, gave us a total storage capacity of 110 tons and a feed rate of about 90 lbs/minute. The system has worked well, and a pH of 7-7.5 has been maintained with about 30 tons/day consumption for the one boiler. This is the equivalent of about 30 lbs/minute of quicklime.

ELRAMA SLUDGE TREATMENT AND DISPOSAL

Under the original concept, it was believed that only fly ash would be removed in the single stage scrubbers. The unfixed sludge would be dewatered in three clay lined ponds, which would be excavated as required, and the sludge placed on our normal landfill. As we and the industry now know, there is always some SO_2 removal, and with the pH we consider necessary, the quantity of SO_2 removed is not negligible. And, as we have sadly learned, unfixed sludge is not easily excavated or transported. For this particular situation and with our recently completed test of the IUCS prototype at Phillips, the POZ-O-TEC system seemed to be the answer to our dilemma. Fortunately, IUCS had just completed their tests at the Mohave Station, and that equipment was available. After entering into a contract with them, the equipment was moved from Nevada, and they were ready to process sludge before we were ready to produce it. Shortly after startup, they were able to obtain and install a vacuum filter which increases the 35-40% solids in our underflow to 50-60%. Consequently, their process requires less dry fly ash to be mixed with the sludge.

COSTS

Thirty-two million dollars has already been spent on the Phillips scrubber system with the expectation of an additional $8 million to be spent for additional equipment to come into compliance with the SO_2 emission limitations for the entire plant. Approximately $34 million has already been spent on the Elrama scrubber system with the expectation of an additional $12 million to be spent for final compliance with the SO_2 emission limitations. These expenses equate to an installation cost of $103 per Kw for the Phillips Power Station and $93 per Kw for the Elrama Station.

The actual operating costs we have been incurring for the Phillips scrubber system, which is not yet complete, are approximately $9 million per year and are equivalent to approximately 5 mills per Kwhr, 38¢ per million BTU, or $8.50 per ton of coal consumed, all including fixed charges. Based on these costs, the combined operating costs for Elrama and Phillips when full SO_2 compliance is achieved will be approximately $32 million per year, or 6 mills per Kwhr, or 52¢ per million BTU, or $12 per ton of coal consumed, all including fixed charges.

The estimated disposal costs of sludge are approximately $6 to $7 per wet ton, or $12 to $14 per dry ton. These costs may change upon selection of a contractor for long term disposal and upon determination of the cost of acquiring and developing additional disposal areas.

We expect that the costs of operating these scrubber systems, when completed, will result in an increase in the customers' bills by approximately 10% over 1974 billing levels.

FUTURE PLANS

Our next phase of construction toward total scrubbing for SO_2 removal at Phillips and Elrama has been known in our

Company as Phase IA. However, with the now recognized importance of continuous pH control, we have gone into an accelerated Phase IA. This is intended to give us redundant lime feed at the Phillips Station by May, 1976, for more reliable scrubber operation, and adequate quicklime feed at the Elrama Station for scrubbing the gas from two boilers by June, 1976.

Completion of the total Phase IA program will provide the lime feed and storage facilities, thickeners, and auxiliary equipment sufficient to enable compliance with SO_2 emission regulations. This will include automatic pH control with the redundancy required for continuous control. This phase should be completed by December, 1977 at both stations. We plan to use magnesium modified quicklime, not only to achieve compliance without the necessity of installing additional vessels, but also to avoid the deposits, which we experience with high calcium lime.

Sludge treatment and disposal is still an unsolved problem in our opinion. Despite glowing reports of presently available methods, we are not convinced. We probably have more experience with the available methods than any other scrubber operator, and none of them has yet been demonstrated as being environmentally acceptable. As mentioned previously, we are in the process of receiving proposals for long term sludge disposal from three different organizations. We have specified that the process chosen must yield a "structurally stable and environmentally acceptable" product. We want all this and a process that we and our customers can afford. As with all scrubber developments, this should prove interesting - and, no doubt, frustrating.

CONCLUSION

In previous papers we had listed what we feel are levels of performance that must be satisfactorily resolved if the system is to be considered operationally feasible. With regard to our experience to date, we presently assess these performance levels as follows:

1. Reliability - Although we have experienced unacceptable reliability with the use of normal high calcium lime at our Phillips Station, our test program using magnesium modified lime indicates that an improved reliability can be achieved. In addition, resolution of several of the problems at Phillips and the implementation of those solutions at the Elrama facility prior to initial operation indicate that some improvement may be achieved in the reliability of the Elrama scrubbers. However, the Elrama system has not yet been run under full load conditions which would more accurately reflect true reliability.

2. Turndown Capability - Over the course of operation in the past year, it has been possible to follow the plant load with little problem provided sufficient scrubber capacity is available. However, we have not been able to determine if such cycling operation contributes to the scaling and deposit problems that have been experienced.

3. Closed Loop Operation - Our goal is to achieve closed loop operation, however, this appears to be several years away. Some improvement was made in reducing blowdown, but it appears that further improvement will have to wait for the use of magnesium oxide lime and the addition of more lime feeders and associated equipment.

4. Sludge Disposal - Our experiences with several sludge disposal techniques indicate that such techniques can result in acceptable stabilization of the sludge. However, there is insufficient information on long term leaching characteristics of this material to claim such techniques as being environmentally acceptable.

5. Cost Consideration - Our installation and operating experiences with the Phillips scrubber system indicate that upon completion of the Elrama and Phillips scrubbers, the operating costs for these two scrubbing systems would increase the customers' bills by approximately 10% over 1974. Although this is not a physical performance level, the economic aspect must be considered in a total system evaluation.

OPERATIONAL STATUS AND PERFORMANCE OF THE COMMONWEALTH EDISON COMPANY WILL COUNTY LIMESTONE SCRUBBER

Warren G. Stober

Commonwealth Edison Company
Chicago, Illinois

ABSTRACT

The Will County Scrubber was installed in 1972 as a full scale demonstration plant. Over the past four years numerous problems have been encountered, many of which have been solved. However, reliable operation on high sulfur coal has not been achieved. This paper reviews the past operating history, some of the major problems and what has been done to solve them, scrubber availability, sludge treatment operation, the cost to own and operate the scrubber system, and future plans for Will County.

OPERATIONAL STATUS AND PERFORMANCE OF THE COMMONWEALTH EDISON COMPANY WILL COUNTY LIMESTONE SCRUBBER

In the spring of 1970, Commonwealth Edison contracted with Bechtel Corporation to investigate the sulfur removal systems then available and to help us select a system having a sufficient degree of development to warrant a large scale installation on our Will County Station's Unit #1.

After deciding upon a wet scrubber system using lime or limestone, a specification was then prepared by Bechtel and released for bid. From the nine bidders that were solicited, seven proposals were received. After detailed study and bid evaluation with special consideration of the project schedule, Babcock and Wilcox was given authorization, in September, 1970, to begin the detailed engineering for a limestone slurry system. The project had a completion deadline of December 31, 1971 which was established by the Illinois Commerce Commission as part of a rate case.

Design

The Babcock and Wilcox designed process was guaranteed to remove 98% of the fly ash and 76% of the sulfur dioxide. These efficiencies were based on a dust inlet loading of 1.355 grains per standard cubic foot at 70°F and burning 4% sulfur Illinois coal.

The Will County wet scrubber was backfitted on a 177,000 gross kilowatt Babcock and Wilcox radiant cyclone boiler that was put in service in 1955. The location of the scrubber and its associated equipment is indicated on Figure 1, Property Plat of Will County Station.

Construction presented a great many problems both physically and schedule-wise. It was necessary to sandwich the scrubber between the boiler house and the service building with a substantial cantilever. Erection of equipment began in May, 1971 and with the use of substantial overtime, the majority of the system was completed by February, 1972. The tight retrofit and quick schedule resulted in a very expensive system: $108 per net kilowatt in 1972 dollars including indirect costs.

The wet scrubber system is divided into three parts: the limestone milling system, the wet scrubber and absorber, and the sludge disposal system.

The milling system as shown on Figure 2 consists of a limestone conveyor, two 260-ton capacity limestone storage silos, two full-sized wet ball mills, each designed to pulverize 12 tons of limestone per hour so that 95% will pass through a 325 mesh screen, and a slurry storage tank. The required limestone is high in calcium carbonate, above 97%, and low in magnesium carbonate, less than 1%.

The wet scrubber is made up of two identical systems, each taking half the boiler flue gas. Each module is designed for 385,000 cfm at a gas temperature of 355°F. Each system consists of two recirculation tanks, venturi and absorber slurry recirculation pumps, a venturi fly ash scrubber, a sump, a two-stage sieve tray sulfur dioxide absorber, a two-stage mist eliminator, flue gas reheater, and an ID booster fan. Each system contains a by-pass damper which permits a module to be taken out of the gas path. Figure 3 illustrates the entire system showing the relative location of each of the above components.

Figure 4 illustrates the flue gas path. The flue gas emerges from the boiler, passes through an existing electrostatic precipitator and on into the venturi. The flue gas velocity at the throat is approximately 130 feet per second with the pressure drop being maintained at approximately nine inches of water. From the venturi the gas flows goes through the sump and then upwards into the absorber. Here the sulfur dioxide is removed as the gas is forced through two separate sieve trays at a greatly reduced velocity, about 10 feet per second. From the absorber the gas passes through a reheater and on to the booster fan.

Figure 5 illustrates the slurry recirculation system. The flow of slurry to each venturi is 5800 gallons per minute. and to each absorber 11,200 gallons per minute. This gives a liquid to gas ratio of about 18 to 1 and 35 to 1 respectively at full load.

The waste slurry is pumped to either a thickener or to back-up settling ponds and all the clarified water runoff is recycled to the scrubber and milling system.

The power requirement to operate the scrubber system is approximately 7,000 kilowatts or about 4% of the unit gross capacity of 177,000 kilowatts. Over and above this power requirement is a need for steam to reheat the gas exiting the system. The system was designed to reheat the gas from 128°F up to 200°F, using about 50,000 pounds per hour of steam at 435°F and 350 PSIG.

Performance 1972-1974

B-module was the first module to come on line on February 23, 1972, and was followed by A-module coming on line on April 7, 1972. Both modules were plagued by more problems than one would expect to be normal during start-up. Demister fouling was very prevalent and numerous changes were made to the demister wash system in an attempt to solve this crippling problem. Other major problems that affected availability were reheater tube vibrations, reheater fouling, scaling on the absorber walls and grid plates, and slurry nozzle pluggage.

Through the end of 1972, A-module operated 1444 hours, the B-module operated 1237 hours, thus attaining availabilities of 29.5% and 25.2% respectively. (Availability, in this case, is calculated by dividing the operating hours of the scrubber by the operating hours of the boiler-turbine unit). Simultaneous operation was 469 hours, for an availability of only 9.6%. The longest period of continuous operation was 21 days, this being with A-module, and the longest simultaneous operation of both modules was 6 days.

By early 1973 many problems had been resolved; however, the major ones remained: demister pluggage, reheater pluggage, and scaling. Along with the recurring major problems, new problems were encountered, such as cracks in the booster fan inlet cones, chloride stress corrosion cracks in the 304 stainless steel reheater tubes and eventual failure of the Corten reheater tubes.

In April 1973 a decision was made to discontinue operation of B-module and concentrate all efforts on achieving satisfactory operation on A-module. If A-module could demonstrate satisfactory operation, then B-module, which originally had a turbulent contact absorber, would be modified so as to be identical to A-module.

A-module, after operating for 1726 hours through November, 1973 (22.6% availability) was taken out of service for extensive changes in the demister, demister wash system, and reheater. The underspray water supply was changed from pond reclaim water to house service water. This change was made to try to prevent the scaling that was taking place on the demister. The new second stage demister was designed to knock out any wash water carryover in the gas flow from the first stage demister top wash system. The intent of these changes was to eliminate demister scaling and carryover into the reheater coils. In an attempt to slow down the rate of deterioration of the reheater coils, a steam supply line was run from the building heating system to the coils for use during scrubber outages. Keeping the coils hot and dry during outages reduced the corrosion due to chlorides deposited on the coils during operation.

A-module returned to service in late March, 1974 following completion of the modifications and an additional delay to repair damage caused by a freeze-up of instruments and the pond recirculation system.

During 1974 we changed the method of calculating scrubber availability from the ratio of scrubber operating hours to boiler operating hours to the ratio of scrubber available hours to the total hours within a given time period. While neither method is completely satisfactory, we believe the new approach provides a more accurate indication of scrubber performance. In some cases this has increased the availability because the repairs to the scrubber were completed before the boiler repairs. However, for the sake of those who still like to look at scrubber availability based on boiler operating hours, these numbers have also been calculated.

In 1974, A-module was available for service 6025 hours for an availability of 68.8%. A-module operated 5468 hours and unit 1 was operational 7,924 hours giving a scrubber availability based on operating hours of 69.0%.

After achieving what seemed like a workable system with A-module, we decided to proceed with the modification of B-module, incorporating all the changes made to A-module. The major changes included the removal of the TCA ping-pong balls and replacement with two stages of sieve trays, the single stage demister being replaced by two stages of a sturdier molded design, and the reheater bundles being replaced with

5/8" OD X 0.065" 316L stainless steel tubes in the lower three banks and similar sized carbon steel tubes in the four upper banks. B-module remained out of service through the end of 1974 and well into 1975.

1975 Operation

We entered 1975 operating in much the same manner as we operated in 1974. A-module was operational and B-module was still undergoing modifications. The first three months of 1975 saw A-module achieve availabilities of 99.5%, 99.4%, and 94.0% (based on hours that the module was available for service). Table I summarizes 1975 availabilities computed by both methods, one based on boiler-turbine operating hours and the other based on the hours the scrubber was available each month.

Figures 7 through 16 are monthly charts indicating all the unit and scrubber outages. While these sheets give the general nature of the outage, they do not specifically mention or detail the work done on the scrubber during unit outages. It should be noted that it was always necessary to do some scrubber maintenance work during those outages.

During the first three months of 1975 there were eight A-module outages. However, only one was a forced outage, the others being for no unit demand, inspections and one accidental trip. The forced outage was due to the splitting of a slurry supply hose to the venturi.

At the time the reasons for the increased availability during 1974 and the early months of 1975 were hard to ascertain. Looking back today at what took place then and what was to follow later in 1975, we believe that the type of coal we were burning at the time had a lot to do with the increased availability. The addition of a second stage demister, automation of pH control and other minor changes had some influence, but were not determining factors. As part of our commitment to reduce SO_2 emissions, only low sulfur (approximately 0.4% S) coal was being shipped to Will County Station. As this type of coal was mixed with and gradually displaced the high sulfur coal remaining in the storage pile, the average sulfur content of coal burned in Unit #1 dropped. In 1974, the average was 1.5%; in early 1975, the average was slightly over 1.0%. We believe that the low sulfur content of the coal and the change from sludge pond reclaim water to house service water for the demister wash were the primary causes of the improved reliability. This conclusion was tentatively confirmed when we attempted operation on high sulfur coal later in 1975 and encountered severe problems.

The change from sludge pond reclaim water to house service water for demister wash water upset the water balance of the system. The high availability of A-module during the first three months of 1975 which resulted in part from this wash water change, caused the water level in our sludge pond to frequently rise to the point where we became concerned about an accidental overboarding to the nearby Des Plaines River. In April the scrubber was shut down until changes could be made that assured overboarding would not occur. Three steps were taken to tighten the water balance and assure a closed loop operation: 1. Pump gland water flows were reduced by 50%; 2. The scrubber house service water filter backwash was routed out of the system; 3. The continuous demister wash underspray was changed to an intermittent spray of 5 minutes on - 5 minutes off. A-module returned to service on May 5, 1975 and the noted changes proved to be effective in maintaining a lower pond level thereby assuring that overboarding would not take place.

B-module returned to service on May 20, 1975 upon completion of its modification program, and the start up was rather uneventful with only minor problems being encountered which resulted in a couple of short outages.

Early in the year it was decided that we should attempt to operate the scrubber on Illinois high sulfur coal. Because Will County was receiving only low sulfur coal as part of our SO_2 compliance program, it was necessary for our Fuel Department to arrange for special shipments of high sulfur coal for the test. Difficulties in obtaining and storing the test coal forced us to begin burning it before we had obtained sufficient experience operating both modules simultaneously or with the tight water balance which had just been put into effect. As a result, we were not totally prepared for the myriad of problems which began to occur.

During the high sulfur coal burn program, SO_2 removal, particulate removal in the scrubber, and precipitator efficiencies were measured along with extensive monitoring of cycle chemistry. Removal efficiency tests were conducted under full and partial load conditions and with the electrostatic precipitator ahead of the scrubber in and out of service. SO_2 removal efficiencies averaged 78.2% on A-module and 86.8% on B-module with inlet SO_2 loadings averaging 3573 ppm. A-module removal efficiencies were lower because only one of the two absorber slurry pumps was in service, resulting in a lower liquid to gas ratio. The other absorber pump was out of service for repair of the impeller and rubber lining.

Particulate removal efficiencies were very erratic and lower than expected. Scrubber outlet loadings, which were measured using the EPA Method #5 train, were high; with a large percentage being slurry carryover rather than fly ash. During the test period slurry solids in the recirculation tanks rose to a higher level than desired. In an attempt to reduce the solids content, additional water was added to the cycle by changing the demister wash cycles. Bottom wash was made continuous and the top wash frequency was increased. The top wash is a deluge system, utilizing a high flow for a short period of time. But since this was pond return water, the increased use of this system prompted scale to form on the walls. Scale and pluggage in the demister apparently resulted in uneven gas distributions and high velocities which permitted slurry to carryover through the system.

Ten days after the start of the test an abnormally high pressure drop was noted across the A-module reheater. An inspection showed the first stage demister starting to plug and the reheater coils heavily fouled. B-module was taken out of service 13 days after the start of the test because of high vibrations of the booster fan. At that time inspection of the module showed severe demister pluggage and the fan rotor coated with a dark scale. Approximately four days were required to clean and balance the fan before B-module could be returned to service.

By mid June, 18 days after commencing the high sulfur coal test program,A-module was out of service due to massive reheater leaks, reheater fouling and demister pluggage. It remained out of service for repairs for the remainder of 1975.

With tight control of the water balance in order to assure a closed water loop, two modules in service and operation on high sulfur coal, changes within the scrubber cycle were occurring quicker than we could detect and correct them. With the turn of events being what they were, it is difficult to assess the operation or draw any firm conclusions from the attempted high sulfur coal test other than to say we were unable to successfully operate on high sulfur coal under these conditions.

Shortly after the test shipment of 20,000 tons of 4% sulfur Illinois coal was consumed, it was decided that coal from the storage pile would be burned in Unit #1. Approximately 90,000 tons of coal ranging from 0.8% to 4% sulfur remained segregated from the remaining 0.4% sulfur coal in storage and was used to fuel Unit #1 when B-module was in service. However, during the months of July, August, and early September 1975 there were occasions when low sulfur (0.4%) coal was used to fuel the boiler and this resulted in a large variation of the inlet SO_2 concentration in the flue gas entering the scrubber. By mid September, B-module was taken out of service because of massive scaling in the absorber, in the sump area just below the absorber and in the demister. Presumably this was caused by the pH control either not functioning properly or the system just not being able to handle such large variations in the inlet SO_2 conditions. An outage of over one week was necessary to hand-chip the scale, which in some places was one-half inch thick, off the various internals.

Another major problem that affected performance of B-module began occurring with less than 1000 hours of operating time on the rebuilt module. This was the failure of the new carbon steel reheater tubes. We eventually experienced failures in six of the twelve new tube bundles over a three month period. Peripheral cracks occurred in the tubes just before they entered the header. In all cases where failures occurred, the tube supports had worked loose permitting the tube to vibrate. The coil manufacturer has repaired and modified all the tube bundles by redesigning the tube supports and their method of attachment to the frame, including an additional support.

Other parts of the system have also experienced problems which are worth noting. The horizontal plate above the nozzles at the venturi inlet wall wash box developed numerous holes allowing slurry to splash and drip over the scrubber structure. The cause was found to be slurry being deflected onto the top plate and eventually eroding through at the point of impingement. Each nozzle has now been fitted with a deflector designed by B&W to redirect the slurry flow. The duct between the absorber outlet and reheater inlet has experienced corrosion to the extent that it has eaten through the ductwork. The ductwork material is ¼" corten steel with a 60 to 80 mil Ceilcote Flakeline 103 coating. After the coating either fell off the wall or was eroded away, it was not long before corrosion took its toll. This area was not the only area in which we experienced problems with coatings. In the sump, the walls between the abosrober were lined with Flakeline and after the scale was removed a thorough inspection of the walls showed areas where the lining was gone and corrosion was taking place. The other half of the sump, that area below the venturi, is lined with 1½ inches of Kaocrete refractory over the Flakeline and appeared to be in excellent condition. The walls below the absorber have since been cleaned and covered with refractory. Severe corrosion of the reheater bundle support frames has taken place. In some areas there is little, if any, of the carbon steel remaining.

One of the absorber recirculation pumps was damaged when a check valve in the pump discharge piping failed and portions fell into the pump. The pump impeller was damaged and its rubber lining destroyed. From the debris found in the pump, it appeared that the rubber lining had eroded on the check, allowing the base material to corrode and ultimately fail. B&W reviewed the need for check valves in the pump discharges and recommend that they be removed. They have since been removed and the spool pieces relined with rubber.

Unit #1 came down for a boiler, turbine and scrubber overhaul in mid October for an outage that lasted more than four months. This gave us an opportunity to conduct an extensive inspection and maintenance program of all scrubber components. Approximately $200,000 was spent during this outage on maintenance material and labor in an attempt to have the scrubber in like-new condition upon its return to service.

Sludge Treatment

There has been little in the way of new development at Will County regarding the sludge treatment and disposal. The treatment methods remain the same as reported in previous meetings, that is, thickener underflow or retrieved ponded sludge is placed in concrete mixing trucks along with fly ash and lime and then transported to a dumping site. Until September 1975, the treated material was dumped into a seven acre on-site clay lined disposal basin where the material would set up after one to two weeks depending on the ambient air conditions. Since September 1975 the Material Service Company, a local quarry and concrete products firm, has been transporting the treated sludge from our facilities directly to a permanent disposal site which they operate.

During 1975, Illinois EPA gave a local group a landfill permit to use the stabilized sludge already in our on-site disposal basin as fill for a future golf course. Over eighty thousand tons of solidified treated sludge have been removed from our basin and transported to the landfill site at a cost to us of $2.35 a ton. There is still additional material in our basin that needs to be removed when weather permits.

Two years ago it was noticed that the solidified treated sludge stockpiled in our on-site disposal basin was crumbling at the surface. Investigation into this phenomenon by Dr. Berger of the Civil Engineering Department, University of Illinois has shown that the treated material loses its ettringite and gypsum properties after undergoing several freeze-thaw cycles. Analysis by x-ray diffraction showed a greater deterioration of these properties as the number of cycles increased. The exact cause of this phenomenon is not totally understood; however, Dr. Berger feels there may be a carbonation of material during the freeze-thaw which lowers the pH, thereby causing deterioration of the ettringite. Perhaps removal of a larger portion of the water from the sludge and compaction of the treated material will reduce the porosity enough such that there is a reduced effect from freeze-thaw cycles.

A belt type vacuum filter has been purchased for Will County and should be in service by mid-summer of this year. Once this filter is in service we will have to change our methods of handling which should reduce the cost of treating the sludge. We also anticipate seeing some improvement in the characteristics of the sludge after the removal of more water. The one big disadvantage to this installation will be the additional water we are going to have in our system which will make it all the more difficult to maintain a closed water loop.

Costs to Own and Operate

Approximate operating costs for the Will County scrubber are shown in Figure 6. These costs are based on actual expenses incurred for operation and maintenance through the year 1975 and are considered approximate because there are a few invoices still outstanding for 1975; however, the numbers appearing in Figure 6 should be considered indicative of the cost to own and operate the scrubber system.

The fact that the scrubber did not have the same availability as the boiler necessitated that the values for the tons of coal consumed, BTU's and net kilowatt hours used in the cost analysis be proportioned to the quantity of flue gas treated by the scrubber when in service.

Included in the cost to own and operate are the following items: annual carrying charges on the invested cost, operating labor, maintenance, limestone, auxiliary power, reheat steam and sludge treatment. Included in the auxiliary power cost is the cost for actual power consumed and the annualized cost of replacement kilowatts for that which is lost to the system because of the scrubber. Reheat steam costs are based on the cost of the fuel required to generate the steam which is removed from the boiler cycle.

The true meaning of maintenance costs is hard to assess when taken over a short time interval. One would not expect maintenance costs to be as high the first year as would be expected after two or three years of operation. Therefore it is essential that a good data base be established over a period of time to get the true picture of what maintenance on these systems actually cost. The cost for maintenance as shown in Figure 6 would have been higher had all the 1975 charges been available for inclusion in the dollar figure shown. Not included in the maintenance cost shown are the costs for the B-module modification. The modification costs were investment dollars rather than expense dollars because equipment of a new design was installed. We anticipate higher maintenance costs in the years ahead when greater operating hours are attained on the system.

The sludge treatment cost includes labor, material and hauling of the treated sludge to the off-site disposal area. Also included is the maintenance cost of the sludge treatment facilities. For the period January through November 27, 741 tons of dry solids scrubber sludge were treated at a cost to Edison of $546,217. When referring to dry solids of scrubber sludge we are talking about the material leaving the scrubber system with no additives or water included. This would be the calcium sulfite and sulfate, unreacted limestone and that fly ash removed in the scrubber's venturi section. On this basis, it cost us $19.69 a ton (dry scrubber solids) to treat and dispose of the material.

To summarize, the cost to own and operate the scrubber was about $22 a ton of coal, 114¢ a million BTU's or 13 mills per net kilowatt hour.

Future Expectations

While the main objective at Will County has been to demonstrate scrubber operation on a full-sized unit, we have, out of necessity, done a considerable amount of experimentation in our efforts to make the system operational. Three main reasons make it impractical to use Will County for R&D projects that would require frequent changes of hardware: (1). Will County Unit #1 while not a base load unit, is an essential part of the Edison system and has had a capacity factor of 50% or greater over the past years, (2). The large size means that any major changes would require large amounts of time and money, and (3). The tight physical configuration puts a constraint on any changes that would require the addition or re-arrangement of equipment.

It has become painfully apparent that operation on high sulfur coal with an assured closed water loop will require considerably more costly experimentation. From past experience, just attempting to keep the system operating will not provide solutions, if any, to the problems at hand. We are in the process of developing a test program to attempt to control, if this is possible, the chemical cycle. The information gained from this program, in addition to possibly improving operations at Will County, may be of some value in designing, building and operating the 450 MW scrubber slated for our Powerton Station.

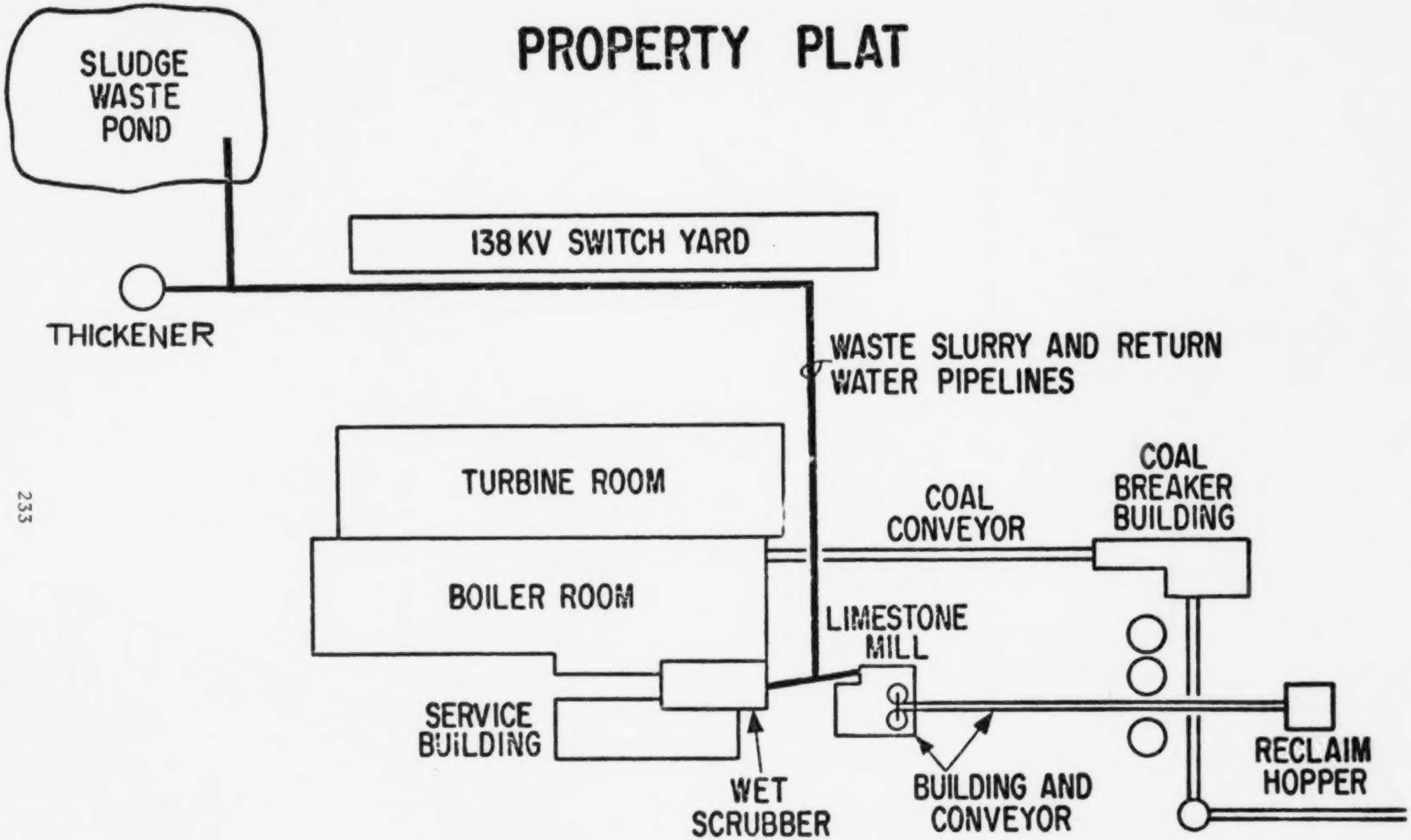
TABLE I

1975 OPERATING HOURS & AVAILABILITY

1975	Generating Unit Operating Hours	Scrubber Operating Hours		Scrubber Availability Based On Unit Operating Hrs.		Scrubber Available Hours		Scrubber Availability Based on The Number Of Hours In Each Month	
		A	B	A	B	A	B	A	B
Jan.	684.5	676	0	98.8%	0	740	0	99.5%	0
Feb.	666	662	0	99.4%	0	667	0	99.4%	0
Mar.	609	604.5	0	99.3%	0	699	0	94.0%	0
April	638	252.5	0	39.6%	0	266.5	0	37.0%	0
May	744	628.5	276	84.5%	37.1%	628.5	276	84.5%	37.1%
June	642	389.5	543	60.7%	84.6%	462	615.5	64.2%	85.5%
July	689	0	547	0	79.3%	0	589.5	0	79.2%
August	564.5	0	568	0	100.0%	0	696	0	93.5%
September	720	0	451.5	0	62.7%	0	451.5	0	62.7%
October*	194.5	0	194	0	99.7%	0	240.5	0	32.3
November *	0	0	0	0	0	0	0	0	0
December *	0	0	0	0	0	0	0	0	0
	6,151.5	3213	2579.5	51.8%	41.9%	3463	2969	39.5%	33.9%

* 21 week scheduled overhaul started October 11, 1975

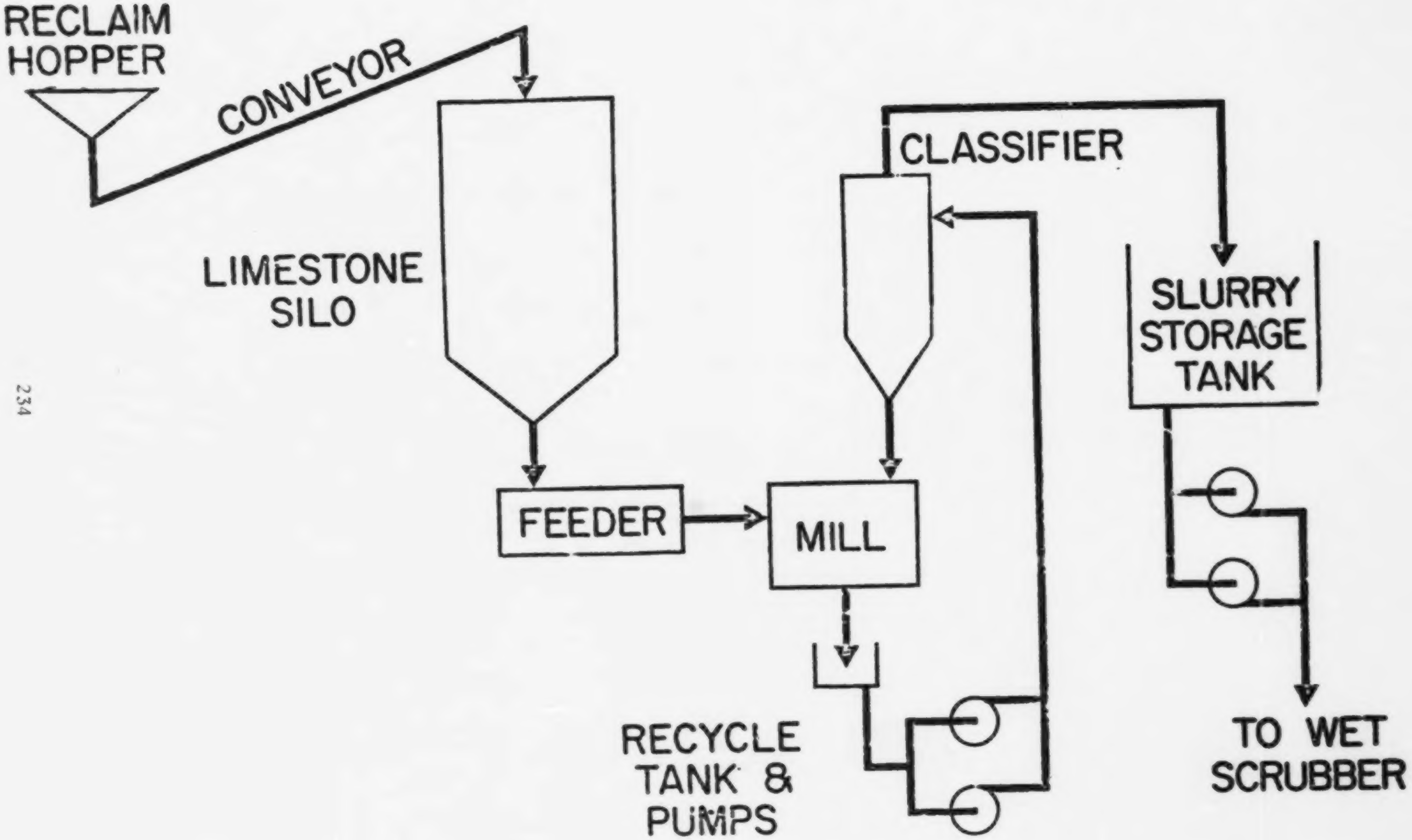
FIGURE 1

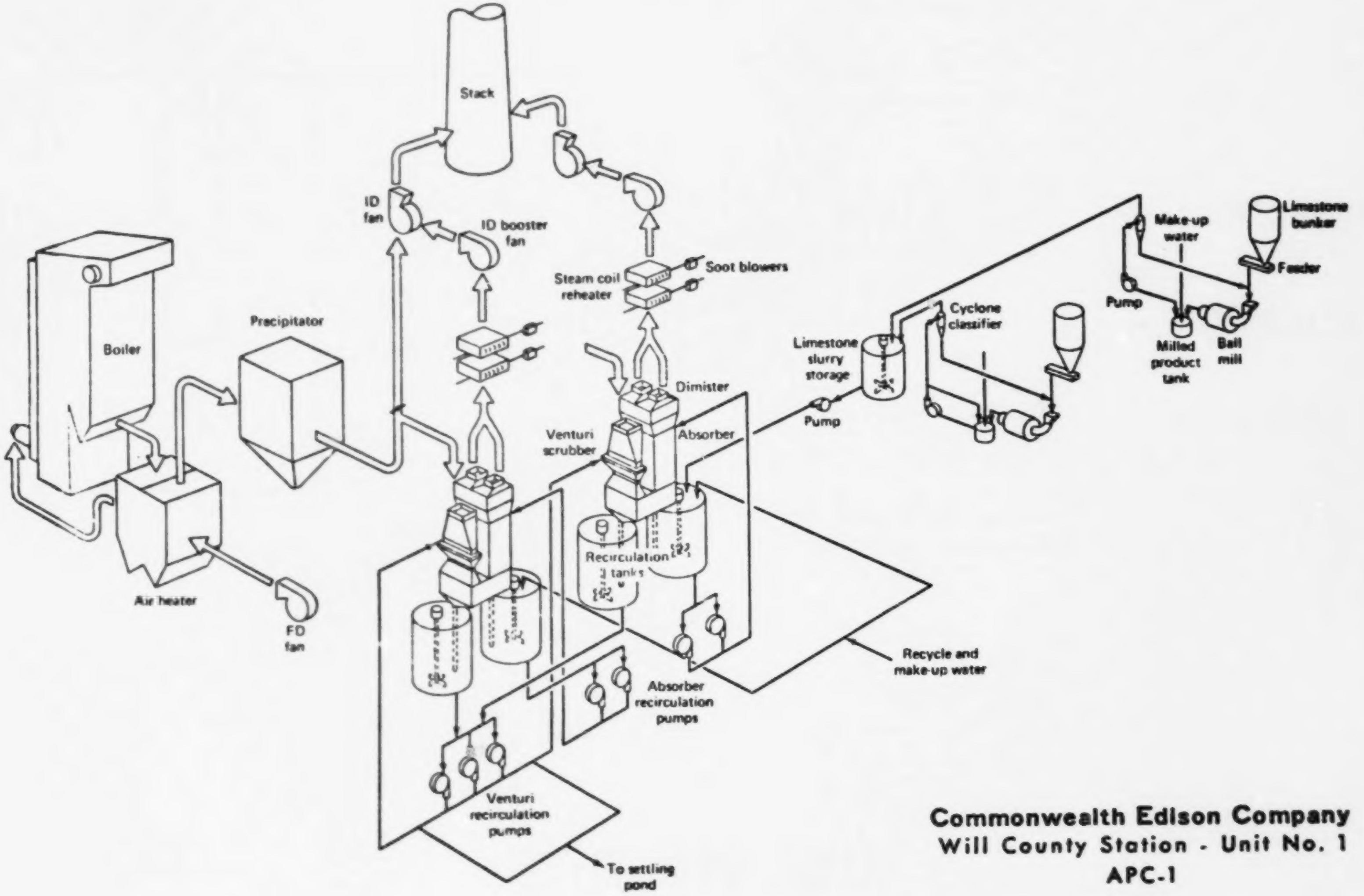


PROPERTY PLAT

SLUDGE WASTE POND
THICKENER
138 KV SWITCH YARD
WASTE SLURRY AND RETURN WATER PIPELINES
TURBINE ROOM
BOILER ROOM
SERVICE BUILDING
WET SCRUBBER
COAL CONVEYOR
COAL BREAKER BUILDING
LIMESTONE MILL
BUILDING AND CONVEYOR
RECLAIM HOPPER

MILLING SYSTEM


RECLAIM HOPPER
CONVEYOR
LIMESTONE SILO
FEEDER
MILL
CLASSIFIER
SLURRY STORAGE TANK
RECYCLE TANK & PUMPS
TO WET SCRUBBER


Stack
ID fan
ID booster fan
Steam coil reheater
Soot blowers
Boiler
Precipitator
Limestone bunker
Feeder
Make-up water
Pump
Milled product tank
Ball mill
Cyclone classifier
Limestone slurry storage
Pump
Dimister
Absorber
Venturi scrubber
Recirculation tanks
Air heater
FD fan
Recycle and make-up water
Absorber recirculation pumps
Venturi recirculation pumps
To settling pond

Commonwealth Edison Company
Will County Station - Unit No. 1
APC-1

FIGURE 3

FIGURE 4

FLUE GAS PATH

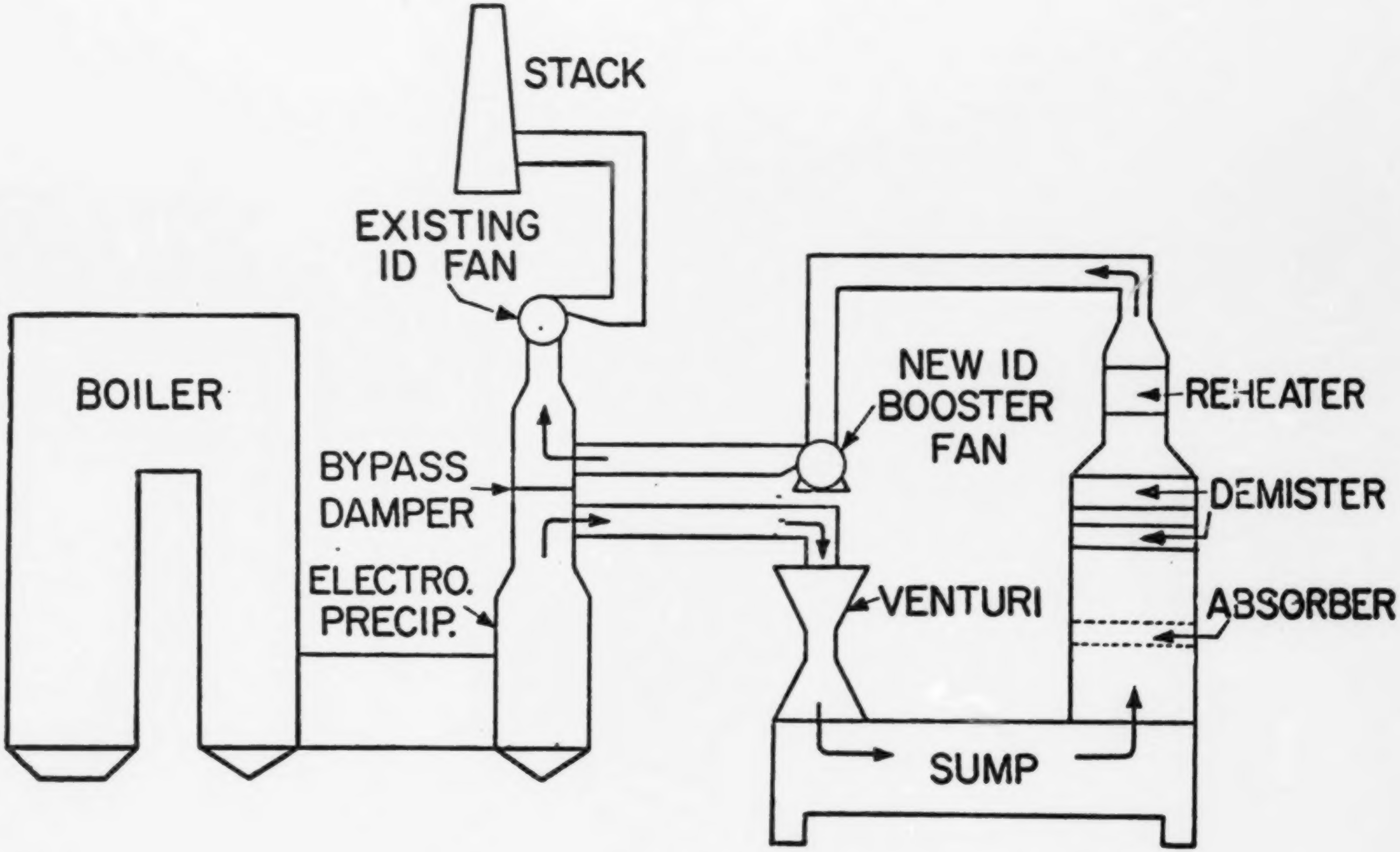

STACK
EXISTING
ID FAN
BOILER
BYPASS
DAMPER
ELECTRO.
PRECIP.
NEW ID
BOOSTER
FAN
VENTURI
REHEATER
DEMISTER
ABSORBER
SUMP

FIGURE 5

SLURRY RECIRCULATION SYSTEM

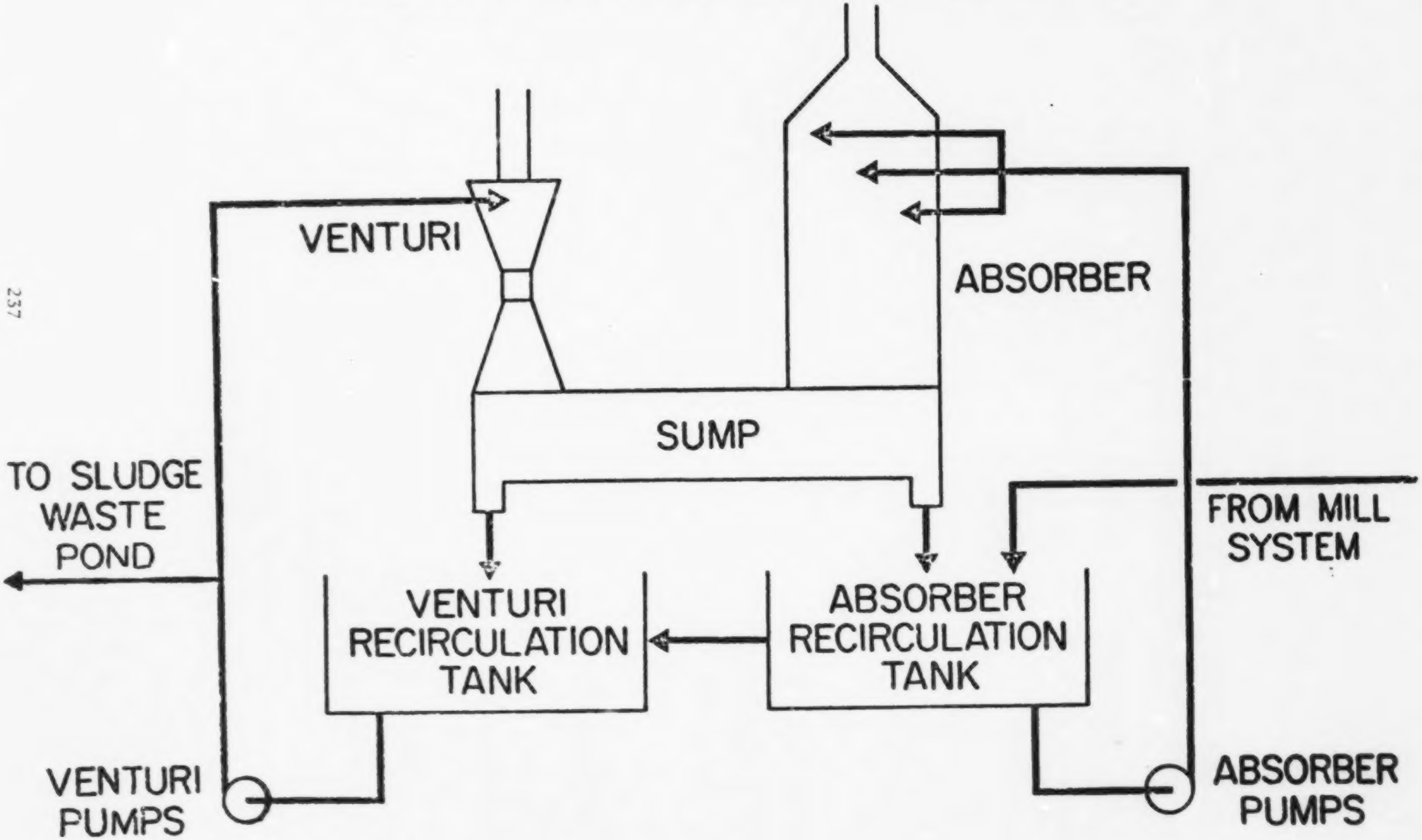

VENTURI
ABSORBER
SUMP
TO SLUDGE WASTE POND
VENTURI RECIRCULATION TANK
ABSORBER RECIRCULATION TANK
FROM MILL SYSTEM
VENTURI PUMPS
ABSORBER PUMPS

FIGURE 6

WILL COUNTY UNIT 1 WET SCRUBBER

APPROXIMATE 1975 COST TO OWN AND OPERATE

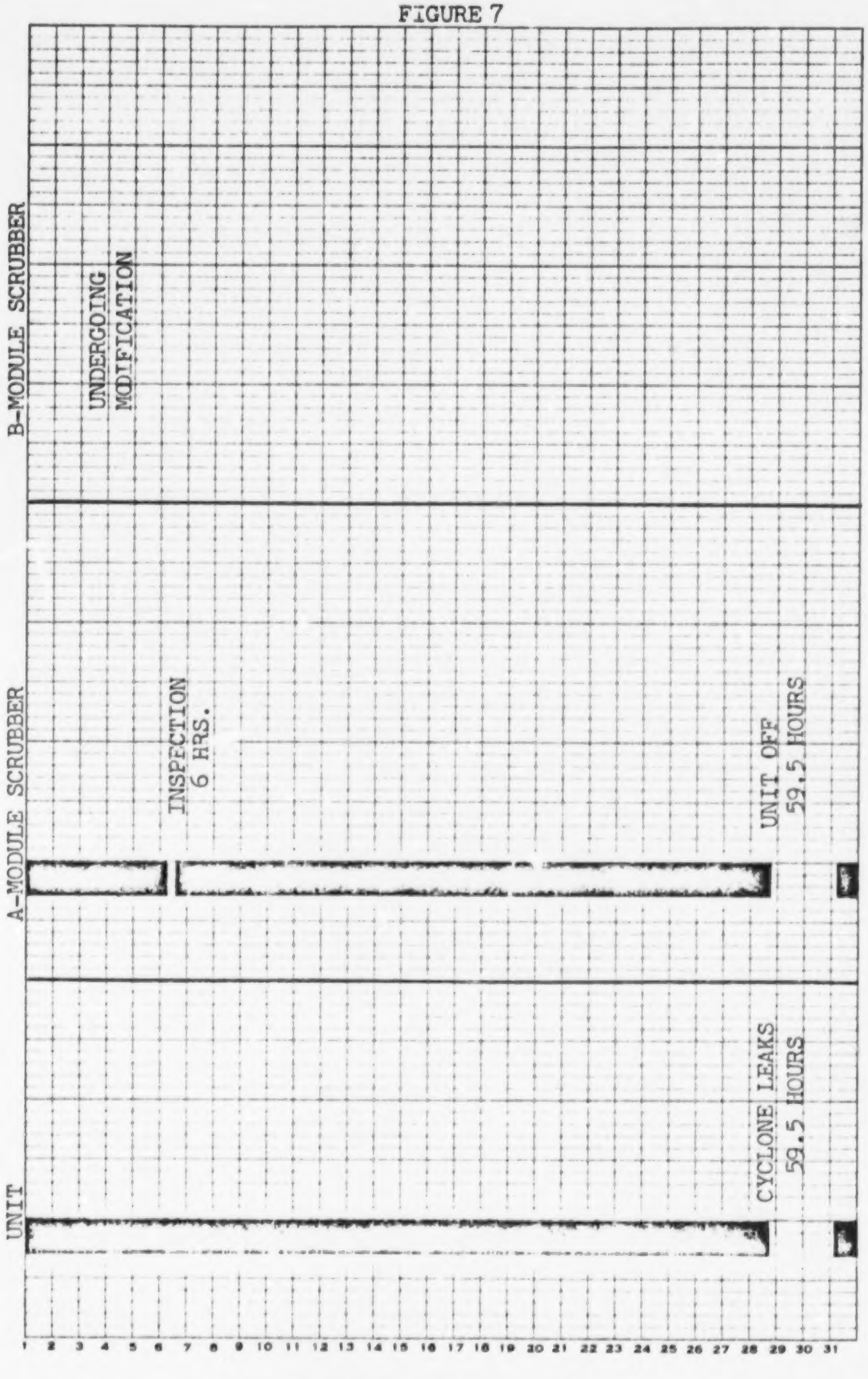
SCRUBBER SYSTEM	$ ANNUAL COST	$/TON OF COAL	¢/MMBTU	MILLS/KWHR
Carrying Charges on $14,900,000	$2,280,000	$ 12.54	65.3 ¢	7.47 Mills
Labor (Operating and Technical)	81,709	0.45	2.3	0.27
Maintenance (Labor and Material)	256,656	1.41	7.4	0.84
Limestone	62,726	0.34	1.8	0.21
Auxiliary Power	586,875	3.23	16.8	1.92
Reheat Steam	72,595	0.40	2.1	0.24
	$3,340,561	$ 18.37	95.7 ¢	10.95 Mills
SLUDGE TREATMENT				
Carrying Charges on $642,000	98.000	0.54	2.8	0.32
Sludge Treatment	546,217	3.00	15.7	1.79
	$ 644,217	$ 3.54	18.5 ¢	2.11 Mills
SCRUBBER & SLUDGE TREATMENT TOTAL COST				
	$3,984,778	$ 21.91	114.2 ¢	13.06 Mills

Notes:

1. Scrubber system has a 14 year life.
2. Sludge treatment cost includes hauling to an off-site disposal area.
3. A-module on line 3213 hours, B-module on line 2580 hours. 181,886 tons of coal consumed, 3.49 X 10^{12} BTU and 305.1 X 10^6 net kilowatt hours are proportioned to scrubber on line hours.
4. Coal average: 9604 BTU/lb, 1.5% sulfur.
5. Unit #1 12 month capacity factor for 1975 was 49.4%.

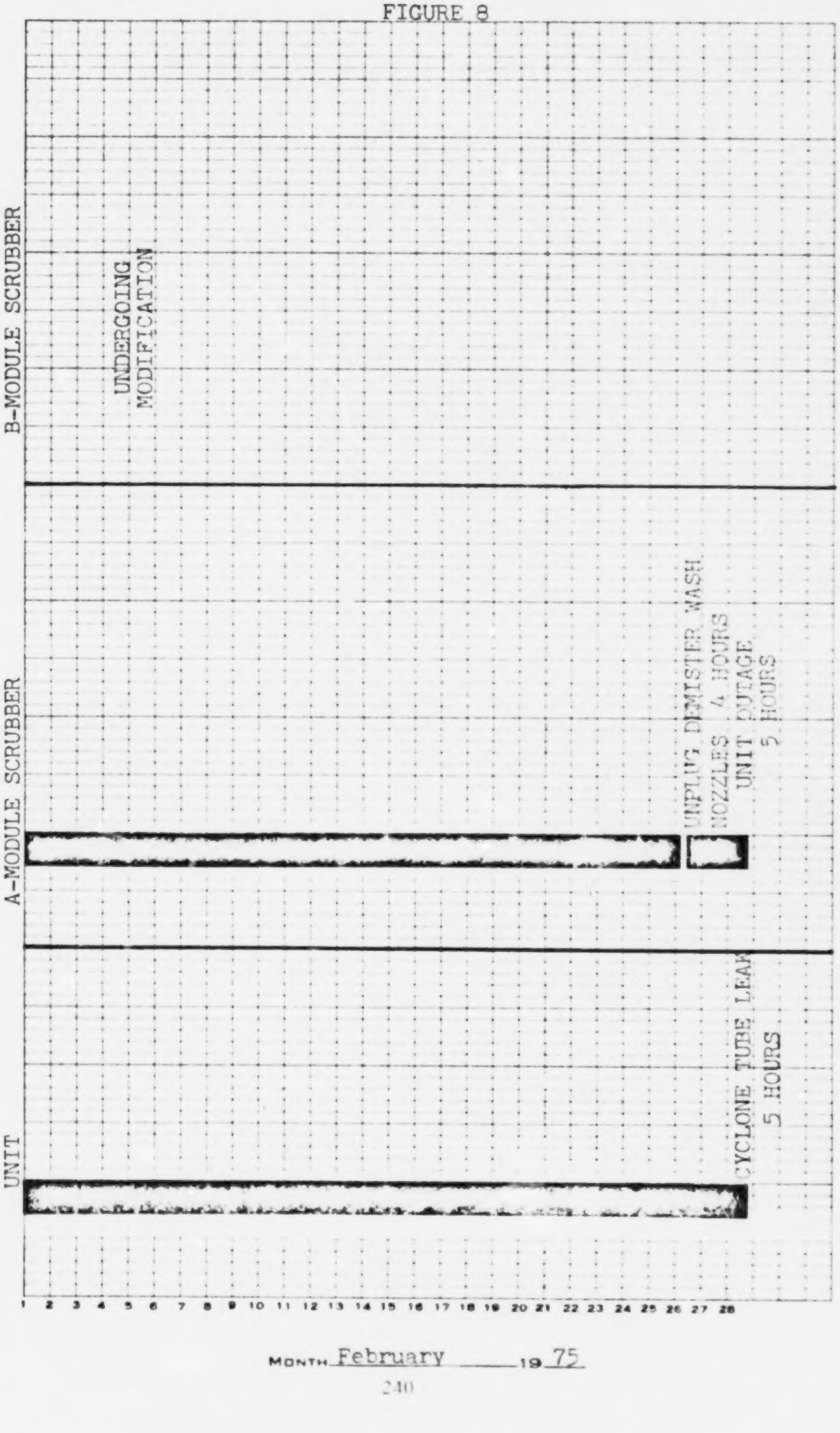
12/30/75

FIGURE 7



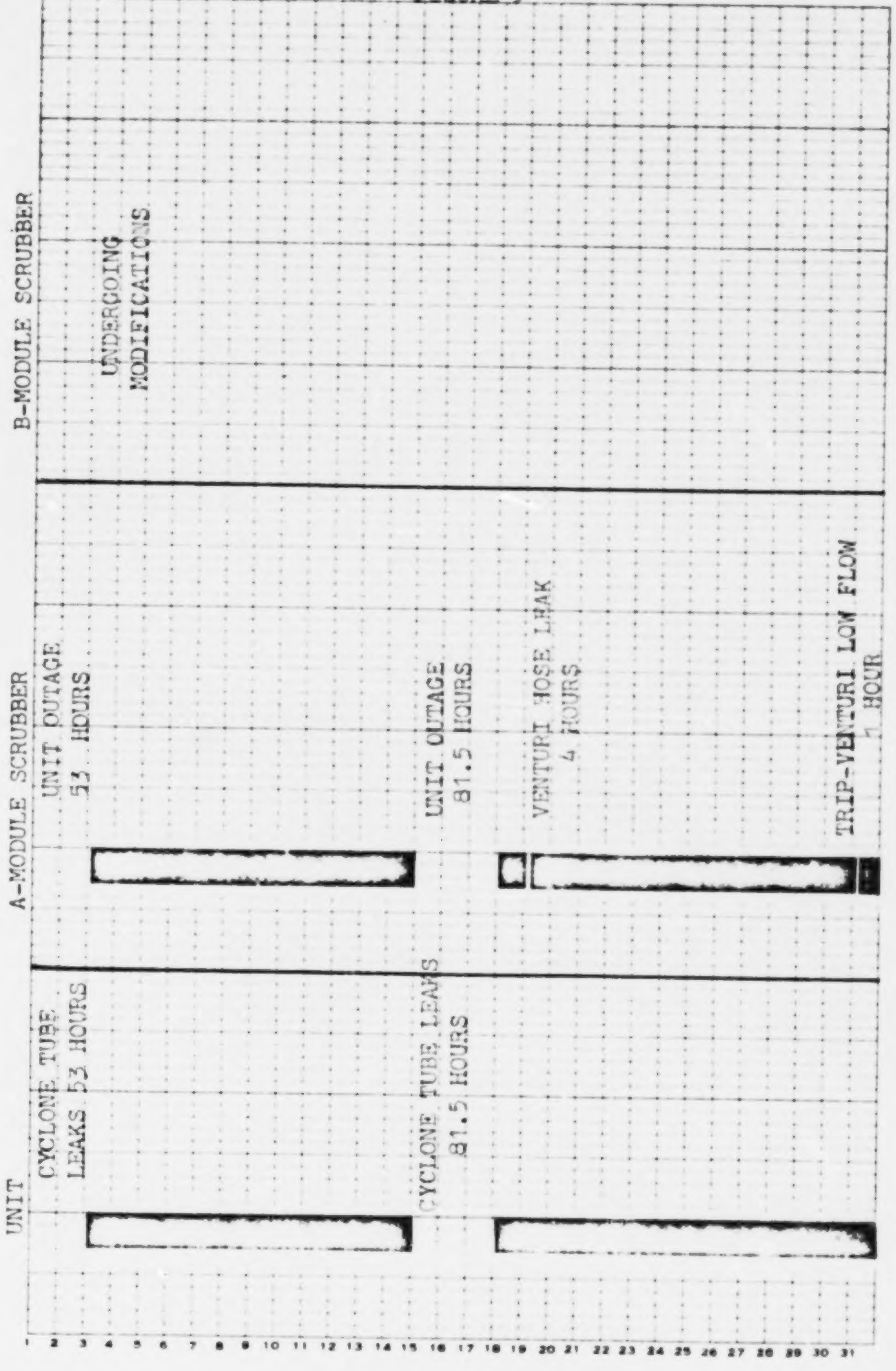
MONTH [illegible] 19 75

FIGURE 8



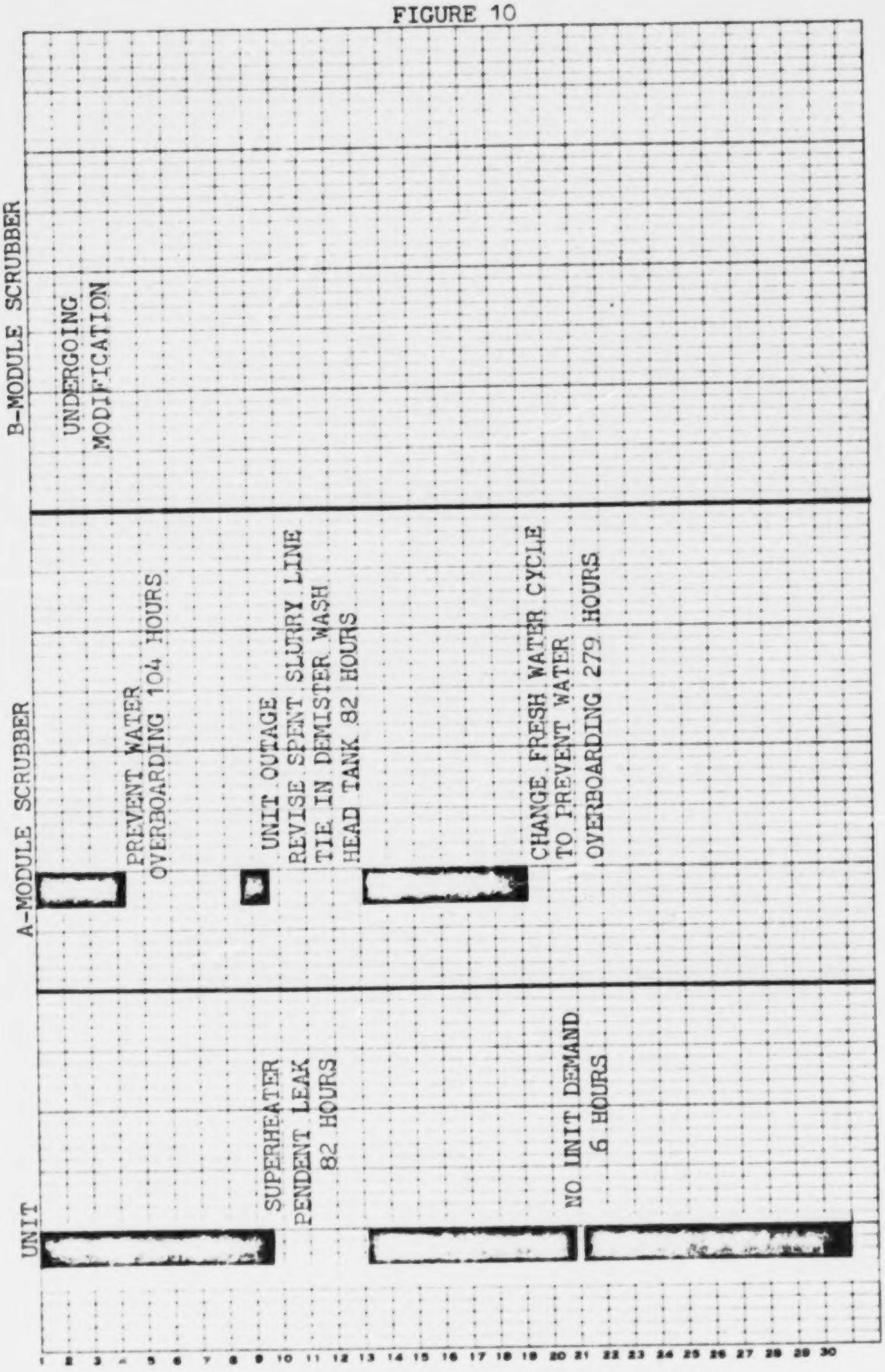
MONTH February 19 75

FIGURE 9



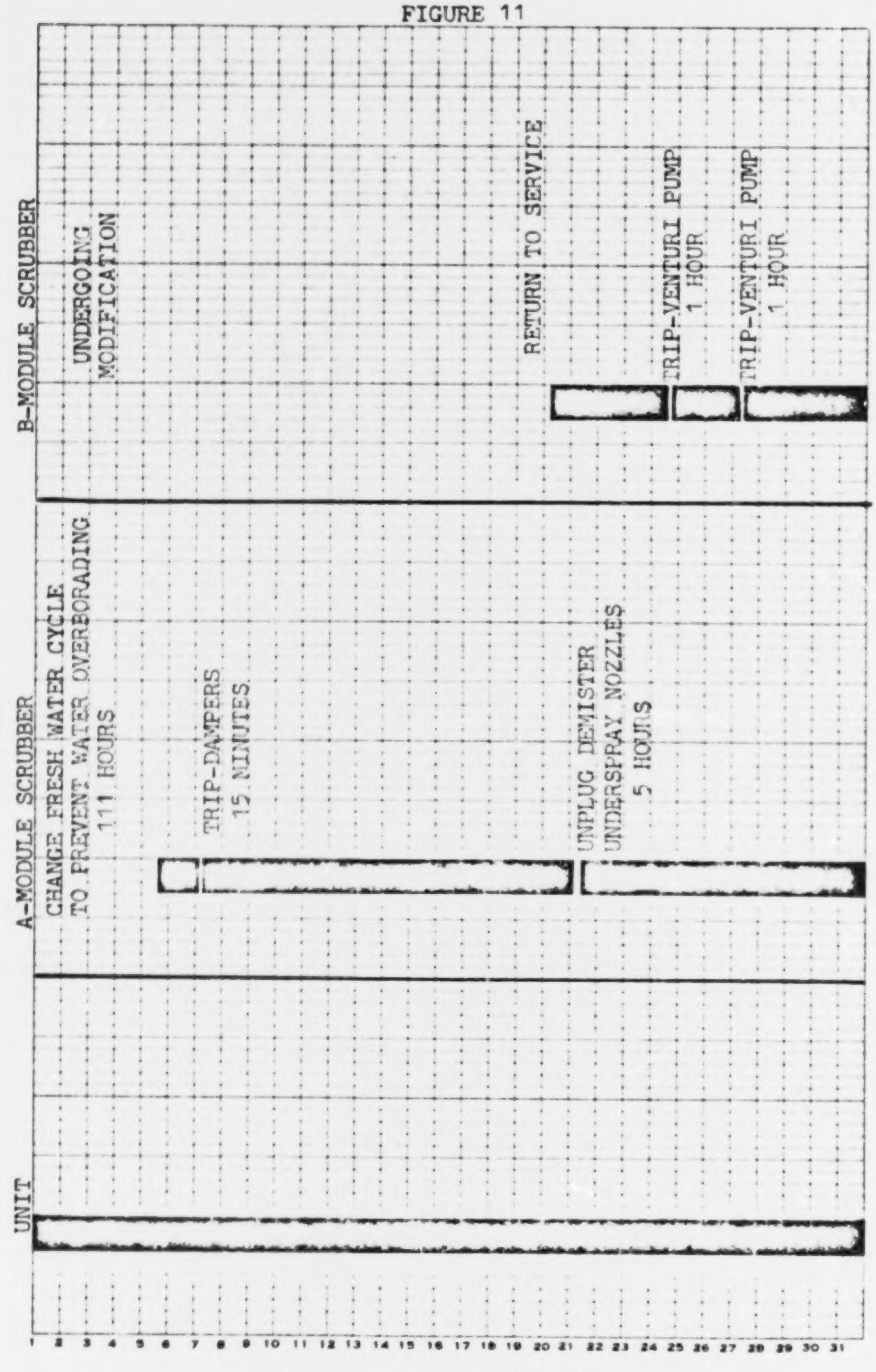
MONTH March 1975

FIGURE 10



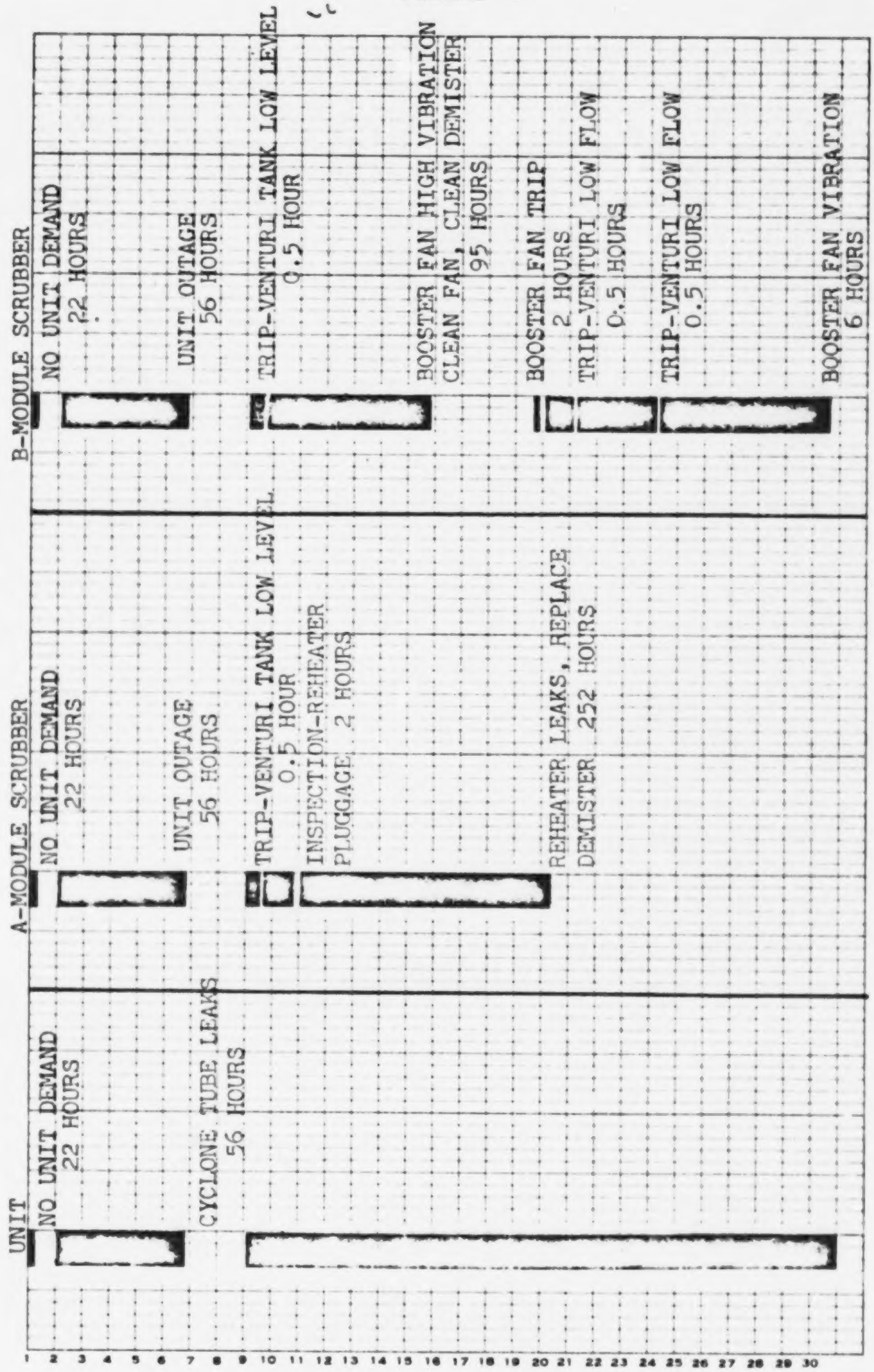
MONTH April 19 75

FIGURE 11



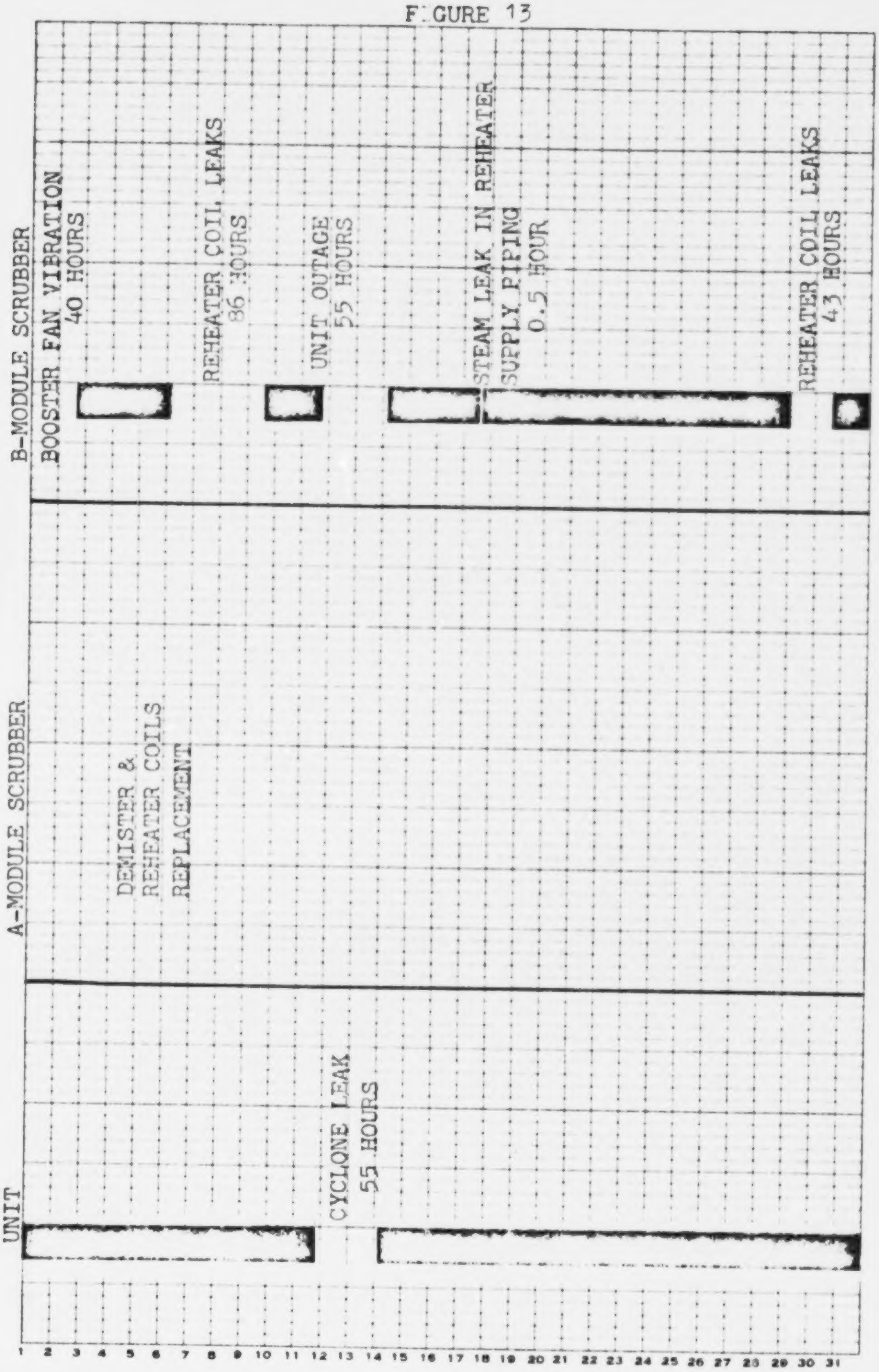
MONTH May 19 75

FIGURE 12



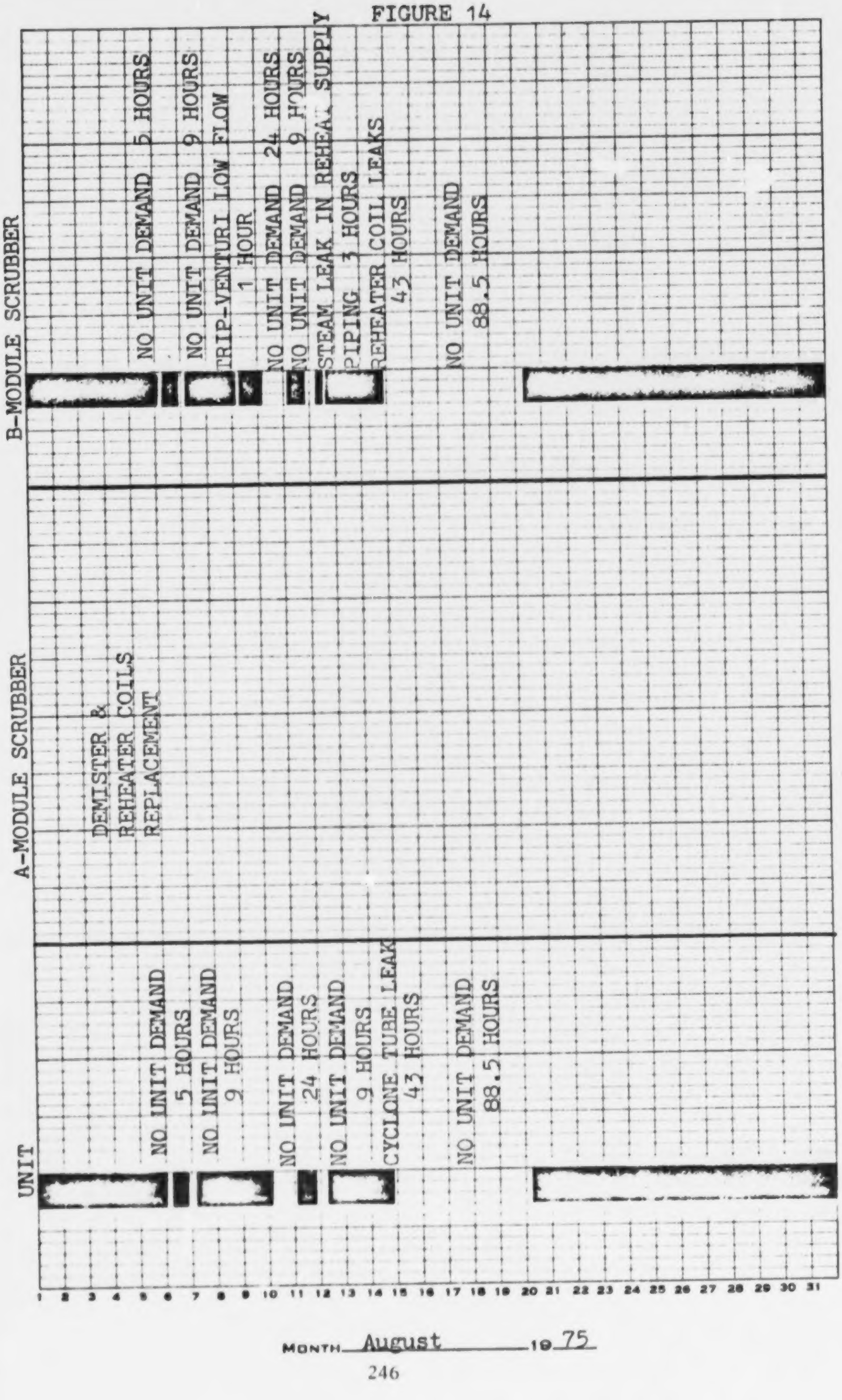
MONTH June 19 75

FIGURE 13



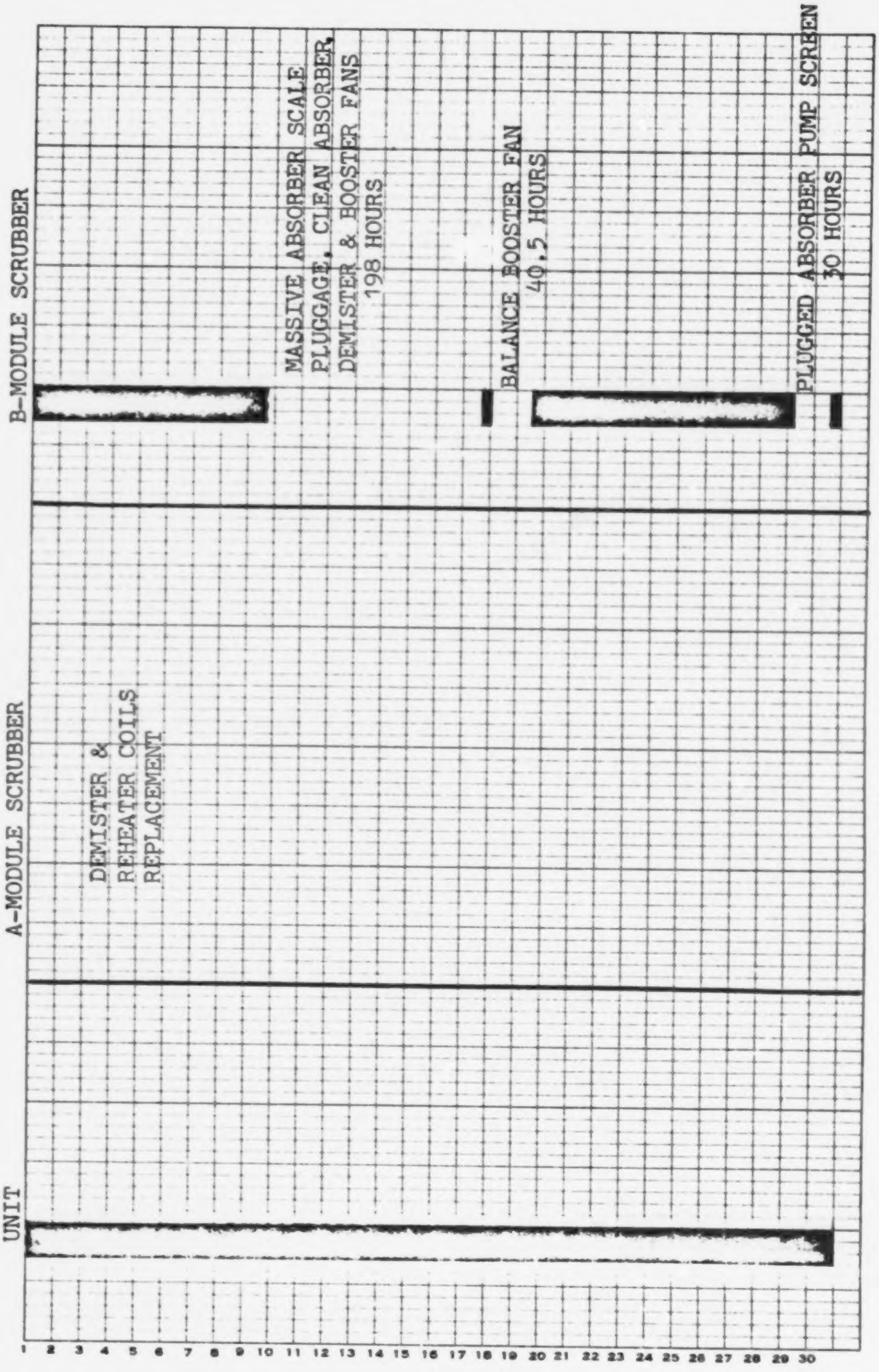
MONTH July 19 75

FIGURE 14



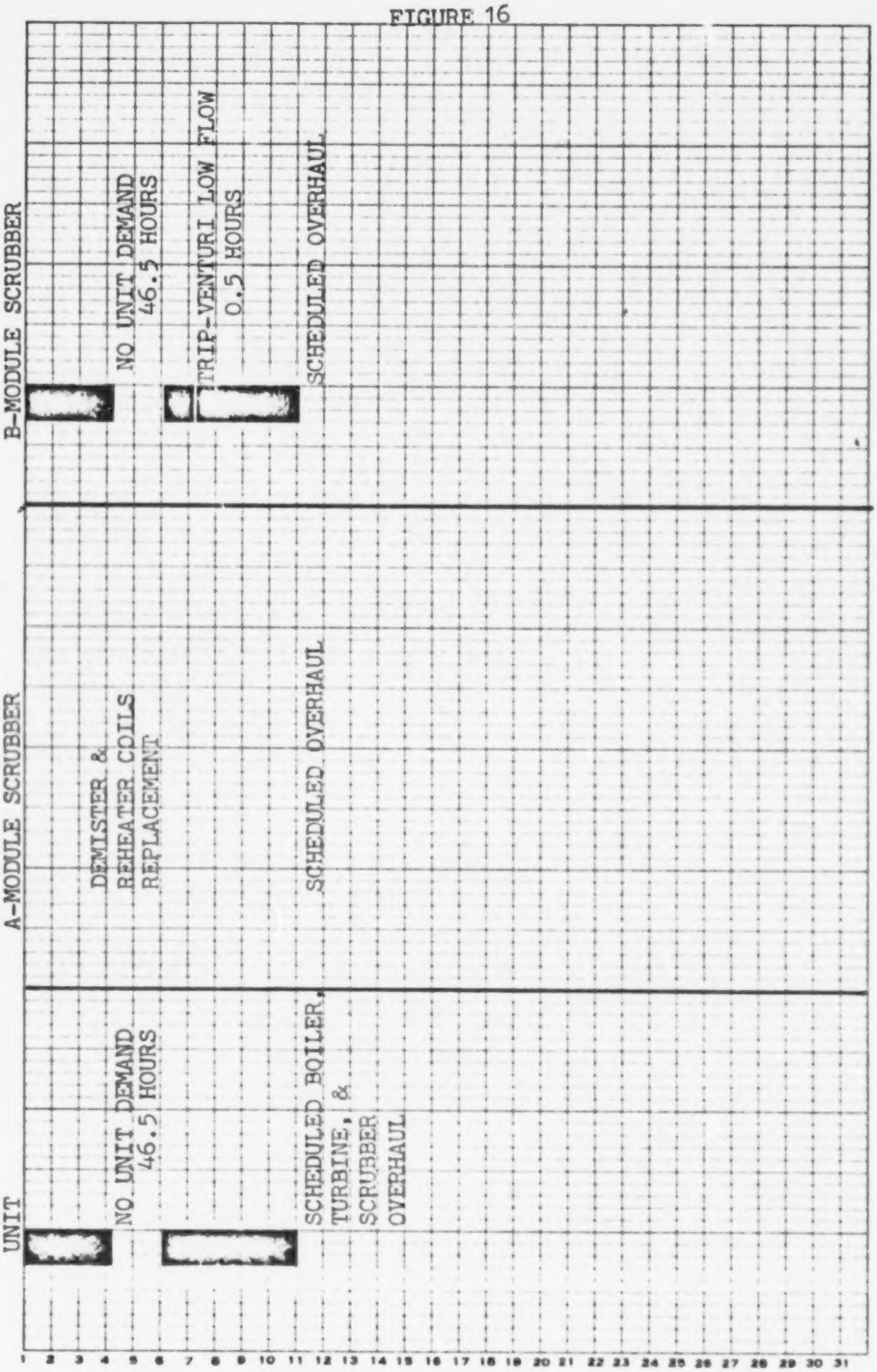
MONTH August 19 75

FIGURE 15



MONTH September 19 75

FIGURE 16



MONTH October 19 75

MHI FLUE GAS DESULFURIZATION SYSTEMS APPLIED TO SEVERAL EMISSION SOURCES

M. Hirai, M. Atsukawa, A. Tatani, and K. Kondo

Mitsubishi Heavy Industries, Ltd.
Tokyo, Japan

ABSTRACT

Since the development of its lime/limestone-gypsum recovery process, Mitsubishi Heavy Industries, Ltd. (MHI) sold thirty-two (32) units of MHI FGD Systems in Japan. Twenty (20) units are operating smoothly, eight (8) are under construction and four (4) are committed by a contract. The total capacity of these installations amount to 11,282,000 scfm, treating gases from boilers, sintering plants and a copper smelter.

This paper describes the features of MHI FGD System, operating experience of its commercial scale units and its applicability to coal-fired boilers.

Further, MHI recently developed a process of removing SO_x and NO_x simultaneously by lime/limestone scrubbing, and its 1,200 scfm pilot test was completed in 1975. Its outline is also reported in this paper.

1. INTRODUCTION

The development of MHI's lime/limestone-gypsum recovery process started 15 years ago, and its intermediate progress report was made public at the Second International Lime/Limestone Wet Scrubbing symposium, back in 1971. (Reference 1)

At that time construction was in progress for a large prototype unit with the capacity to treat 20% of the flue gas from a 156 MW power unit. This prototype plant started up in March, 1972 and as its operation was very satisfactory (Reference 2), MHI has been successful in marketing its systems in Japan. When limited to utilities, MHI's share in the market is about 56% in terms of treated gas flow. (Reference 3)

MHI's system development is shown in Figure 1, and its delivery record is listed in Table 1.

Among the 20 units in operation, 15 units are for oil-fired boilers, 4 units are for sintering plants (steel mills) and the remaining one unit is for a copper smelter. The capacity of most of these installations ranges from 250,000 to 740,000 scfm in terms of gas volume treated, and among the units so far installed, the largest capacity of one module of a scrubber is 560,000 scfm. SO_2 concentration in the flue gas ranges from 500 to 23,000 ppm. Lime or Limestone is used as absobent and high quality gypsum is recovered as a by-product which is used for cement and wallboard production.

The prototype plant mentioned above has so far been operated for about 4 years, and it is now being operated as a 156 MW full-scale plant, in combination with a plant which treats the remaining 80% of the flue gas. Further, another 156 MW full-scale plant is under construction in the same power station.

In evaluating a lime/limestone scrubbing system, the following three items are considered to be important.

1) Reliability

2) Economics

3) Sludge disposal

From the start point, MHI's aim were to fulfill the above requirements and so far have been quite successful.

This paper describes the features of the MHI Lime/Limestone-Gypsum Process, operating experience of its commercial scale system and its applicability to coal-fired boilers.

Recently, NO_x removal has become a serious problem. For several years, MHI has been conducting a series of R & D programs on both wet and dry NO_x removal processes. One of the successful developments is a process of removing NO_x and SO_x simultaneously in a lime/limestone scrubber at high efficiency, by use of a water soluble catalyst. As its pilot test was completed in 1975, its outline is introduced in this paper.

2. SYSTEM DESCRIPTION

The MHI FGD System is a wet non-regenerative system which uses lime or limestone as absorbent and recovers high quality gypsum as by-product. A simplified flow scheme of the system is shown in Figure 2.

The flue gas is fed into the precooler to be cooled down, most of its particulates being removed there at the same time before entering the SO_2 scrubber. The absorbent slurry for SO_2 absorption, and the gypsum "seed" slurry for scale prevention, are fed to the scrubber. The cleaned flue gas passes through the mist eliminator and is then reheated to restore buoyancy.

The scrubber slurry flows into the sulfite oxidizer where all the calcium sulfite in the slurry is oxidized by air, yielding gypsum slurry. A portion of the gypsum slurry is recycled to the scrubber and the remainder is transferred to the gypsum recovery section of the system. Here, the gypsum slurry is concentrated by a thickener and the underflow slurry is fed to centrifuges, yielding by-product gypsum. Most of the thickener overflow is reused for absorbent preparation. Blowdown water from the precooler and part of the thickener overflow are fed to the water treatment section of the system.

Each constituent of the MHI FGD System has the following objectives and features.

2.1 Precooler

Main objectives are as follows:

1) Removal of dust from the incoming gas to obtain high quality gypsum.

2) Cooling of the flue gas to protect plastic materials used in the SO_2 scrubber.

3) Humidification of the flue gas to avoid any local evaporation which may lead to deposit formation in the SO_2 scrubber.

A spray tower is used to economically achieve the above requirements. The spray pump capacity in terms of L/G is normally 14 and gas stream pressure drop (including demisters) is about 1 1/2 in.

2.2 SO_2 Scrubber

Important requirements of this equipment are as follows:

1) Stable and high SO_2 removal efficiency even at severe load fluctuations.

2) Scale-free operation.

3) Low pressure drop (i. e. low power requirement).

4) Low mist entrainment at the outlet.

In order to accomplish the above requirements, MHI has developed a packed spray tower employing low density "grid" packings. Simple structure of the "grid" packings enables the scrubbing slurry to be distributed uniformly on its surface, thus, giving a large absorption area. Another feature of the "grid" packed tower is its low mist entrainment at the outlet which contributes to the reliability of the mist eliminator. The scrubbing slurry pumping rate in terms of L/G is 50 to 140.

Scale-free operation is accomplished by using relatively high L/G and adding gypsum "seed" slurry to the scrubber. The gas stream pressure drop is about 1 1/2 in.

2. 3 Mist Eliminator

The mist must be removed from the gas at high efficiency without scaling of the eliminator surfaces. All of the MHI FGD Systems that are now in operation are using vertical chevrons, giving very good service. Pressure drop is within the range of 1/2 to 1 in.

2. 4 Reheater

Reheating the scrubber gas is necessary to restore buoyancy, to reduce visible steam plume, and to protect downstream equipment from possible corrosion.

2. 5 Absorbent Preparation

This section of the process is important bacause it gives great influence on the following.

1) SO_2 removal efficiency and absorbent utilization.

2) By-product gypsum quality.

In order to achieve this, the ultimate method is to make the particle size of the absorbent slurry as small as possible.

2.6 Sulfite Oxidizer

The oxidation reaction of calcium sulfite to calcium sulfate depends on dissolution and diffusion of oxygen into the slurry. MHI is using a rotary atomizer, an air atomizer of unique structure, which generates fine air bubbles at operating pressure of about 50 psig. The oxidation rate is also improved by lowering the slurry pH. A small amount of sulfuric acid is fed to the oxidizer for this purpose.

2.7 Gypsum Recovery

Gypsum crystals produced in the oxidizer are relatively large, blocky crystals that settle easily. Therefore, the gypsum slurry can be concentrated in a reasonable size thickener and dehydrated by conventional centrifuges. The by-product gypsum contains only 5 to 10% of free water. Gypsum is a stable, harmless compound that can be used for wallboard production and as a retarder for cement setting. In the case of over-supply, the by-product gypsum is more suitable for landfill disposal than sulfite sludge.

2.8 Water Treatment

Some water is disposed of to keep the concentration of chloride and other impurities in the system at a safe level.

Bleed streams from the precooler and the thickener overflow are treated to meet the local water pollution regulations.

2.9 System Control

The controls for the MHI FGD System depend upon the type of absorbent used. In lime scrubbing, the slaked lime slurry feed is controlled by the pH of the recirculating scrubber slurry. In limestone scrubbing, the limestone slurry feed is controlled by the SO_2 in the flue gas and the pH of the recirculating scrubber slurry.

3. OPERATING EXPERIENCE

MHI has experience in construction and operation of the FGD systems applied to several emission sources. They are oil-fired boilers, sintering plants (steel mills) and a copper smelter.

Each emission source has different flue gas conditions, and consequently, the associated FGD system runs somewhat differently from one another. Operating data obtained from the commercial plant operation are described in Table 2.

Operating experiences in the commercial plants, problems encountered, and solutions to these problems are outlined below.

3.1 Precooler Operation

Problems on materials of construction were experienced due to low-pH (1 to 1.5) and corrosive nature of the recirculating spray water.

1) Pin-holes on rubber linings due to bad workmanship caused corrosion on carbon steel underneath.

   This problem was overcome by setting stringent standards for lining works and pin-hole inspection.

2) In an installation for a sintering plant, the incoming gas laden with oily matter caused swelling of rubber linings.

   This problem was solved by using oil resistant rubber linings.

3) In an installation which used sea water to prehumidify the flue gas, accumulated chloride ions in the recirculating spray water caused corrosion on stainless steel valves. The problem was solved by using rubber lined valves.

3.2 Scrubber Operation

3.2.1 SO_2 removal efficiency and absorbent utilization

Independent of flue gas conditions and types of absorbent used, the SO_2 concentration at scrubber outlet ranges from 10 to 60 ppm, or more than 90% in terms of removal efficiency.

Past operating data shows that the absorbent utilization ranges from 90 to 99% or on the average of 95%. One of the reasons why the absorbent utilization is high even in limestone scrubbing is that most of the limestone in Japan is of high calcium in nature.

3.2.2 Scrubber control

Using the pH of the recirculating scrubber slurry as a control factor, all of the scrubbers are running with excellent stability. For example, the SO_2 concentration of flue gases from sintering plants changes from 800 to 1,200 ppm every 20 minutes. Even at this kind of fluctuation, the FGD system operated with high and stable SO_2 removal.

3.2.3 Scrubber scaling

One of the important requirements in slurry scrubbing is scale-free operation of scrubbers. In the early stage of development, some scaling problems were experienced due to poor operating procedures. According to MHI's experience, the following considerations are vital in preventing scaling.

1) Addition of gypsum "seed" crystals to provide seed sites for calcium sulfate crystallization.

2) Application of relatively high L/G and selection of a scrubber with simple structure, to maintain chemical conditions of the scrubbing slurry uniform, and to keep all of the internals wet and clean.

The above method for preventing scaling was proved to be a good way, even in units for sintering plants where the sulfite oxidation in the scrubber reaches as high as 100%, due to high O_2 concentration of the incoming flue gas.

3.3 Mist Separation

The scrubber outlet mist loading ranges from 4 to 8 gr/acf.

Using vertical chevrons, more than 99% of the entrained mist is separated from the gas. Fresh water and a part of the scrubbing slurry is used for washing the eliminator surfaces. Continuous washing and intense intermittent washing were experienced, both giving excellent service.

3.4 Gas Reheating

Most of the installations reheat the scrubber gas directly by burning low-sulfur fuel oil. Installations for sintering plants burn combustible flue gases from steel mills.

One installation for small industrial boilers do not reheat at all. The reheat temperature is rather high in Japan, usually in the range of 250 to 285 °F.

3.5 Absorbent Preparation

Both for lime and limestone, recycle water is used for preparation of the absorbent slurry. In lime slaking, the grits are ground in a wet mill and carried along with the slaked lime into the scrubber. Limestone is ground in a wet mill, but recent installations are using fine limestone powder.

3.6 Lime vs. Limestone

Recent installations tend to use limestone rather than lime. According to MHI's experience, the difference between limestone and lime can be outlined as follows.

1) Lime is more reactive, and therefore lime scrubbing is much easier to control.

2) Sulfite oxidation rate is much higher in limestone scrubbing.

3) Absorbent preparation is much easier with limestone powder.

   In lime slaking, the substitution reaction between lime and the dissolved magnesium sulfate in recycled water may cause calcium sulfate scaling. Here again, the addition of gypsum "seed" crystals in the slaker is very effective for scale prevention.

4) Limestone is more economical, safer and easier to handle.

3.7 System Reliability

All of the installations are running with high availability. Availability data on some installations are shown on Table 3.

4. APPLICATION ON COAL-FIRED BOILERS

Since most of the utilities in Japan are using oil in their boilers, all of the MHI FGD System installations for boilers are treating oil-fired gases. If an electrostatic precipitator is installed before the system, main differences between oil and coal-burnt gases would be in compositions of dust and gas. MHI has been conducting field pilot tests on a coal-fired boiler in Japan. Test results obtained up to now indicates that the MHI gypsum recovery process is applicable on coal-fired boilers without problem. On the other hand, applicability of the system to non-recovery (throwaway) processes has been tested also. If a non-recovery process is favorable, which is the case in the United States (Reference 4), the MHI FGD System can meet this requirement, and can be simplified as follows.

1) Precooler can be eliminated.

2) Absorbent preparation section can be simplified.

3) Gypsum recovery section can be eliminated.

4) Sulfite oxidizer capacity is minimized to gypsum "seed" production.

For application in the United States, tests on dolomitic lime/limestone scrubbing are also in progress. Through these tests, influences of chloride and other impurities on the system have been investigated. Test results up to now indicates that the system is quite feasible. Moreover, studies on materials of construction, process equipment, process economics are being continued for further development of the system.

5. MHI SIMULTANEOUS NO_x/SO_x REMOVAL PROCESS

MHI's research for the NO_x removal process has been aimed at the development of reasonable and economical systems which can treat flue gases ranging from clean flue gases of LNG-firing boilers to dirty flue gases of high sulfur fuel-firing boilers, sintering plants (steel mills), etc. MHI has developed both dry and wet processes of NO_x removal and has already obtained orders for the NO_x removal units. For dirty flue gases requiring both NO_x and SO_x removal, MHI is considering that the use of the wet process is most reasonable for them, in view of its capability of simultaneous NO_x/SO_x removal in a single scrubber and from the standpoints of economics, operation and maintenance, installation area, etc.

Recently, MHI has successfully finished its preliminary stage of development for a simultaneous NO_x/SO_x removal process. By adding an O_3 generator and an ammonia recovery section to the MHI FGD System, NO_x and SO_x in the flue gas are removed simultaneously in a single scrubber, at high efficiencies. The scrubber is identical to those used in the MHI FGD System, lime or limestone can be used as absorbent, and by-products are gypsum and ammonia.

5.1 Process Development

After a bench scale test (125 scfm capacity) for about a year, a pilot plant test (1,250 scfm capacity) was conducted from July, 1975. The flue gas from an oil-fired boiler contained 180 to 220 ppm NO_x, 800 to 1,100 ppm SO_x, and 0.04 to 0.05 gr/scf dust. The obtained efficiencies were 80 to 90% NO_x removal and over 95% SO_x removal. Since the result was satisfactory, a prototype scale unit is planned in the near future.

5.2 Process Description

A simplified flow scheme of the process is shown in Figure 3.

The flue gas is first cooled and dust collected in the precooler and after the oxidation of NO to NO_2 by means of adding stoichiometric amount of O_3, NO_2 and SO_2 are removed simultaneously in the SO_2/NO_2 scrubber. The absorbent slurry, either slaked lime slurry or limestone slurry, containing water soluble catalyst is fed to the scrubber.

Since the scrubber outlet slurry contains the N-S compounds (defined as a mixture of nitrogen and sulfur base compounds) as dissolved components, a portion of filtrate from the gypsum recovery section is sent to the ammonia recovery section where it is decomposed to yield gypsum and ammonia.

Description of the remaining section of the process is identical to those of the MHI FGD System.

5.3 Features

The outstanding features of the process can be summarized as follows:

1) Most of the main equipment are exactly the same as that of the MHI FGD System and their reliabilities are already known to be high through actual operation.

2) Since NO_x and SO_x are removed simultaneously in one system, the process has advantages in installation area, operation and maintenance, economics, etc.

3) As the reductive property of SO_2 is used in absorption, the process does not require any NO_x reducing agent. (In the dry process, ammonia is added as the reducing agent)

4) The process is a recovery system which produces gypsum and ammonia as by-product.

5) The process can be applied to the existing MHI FGD System by simple reconstruction.

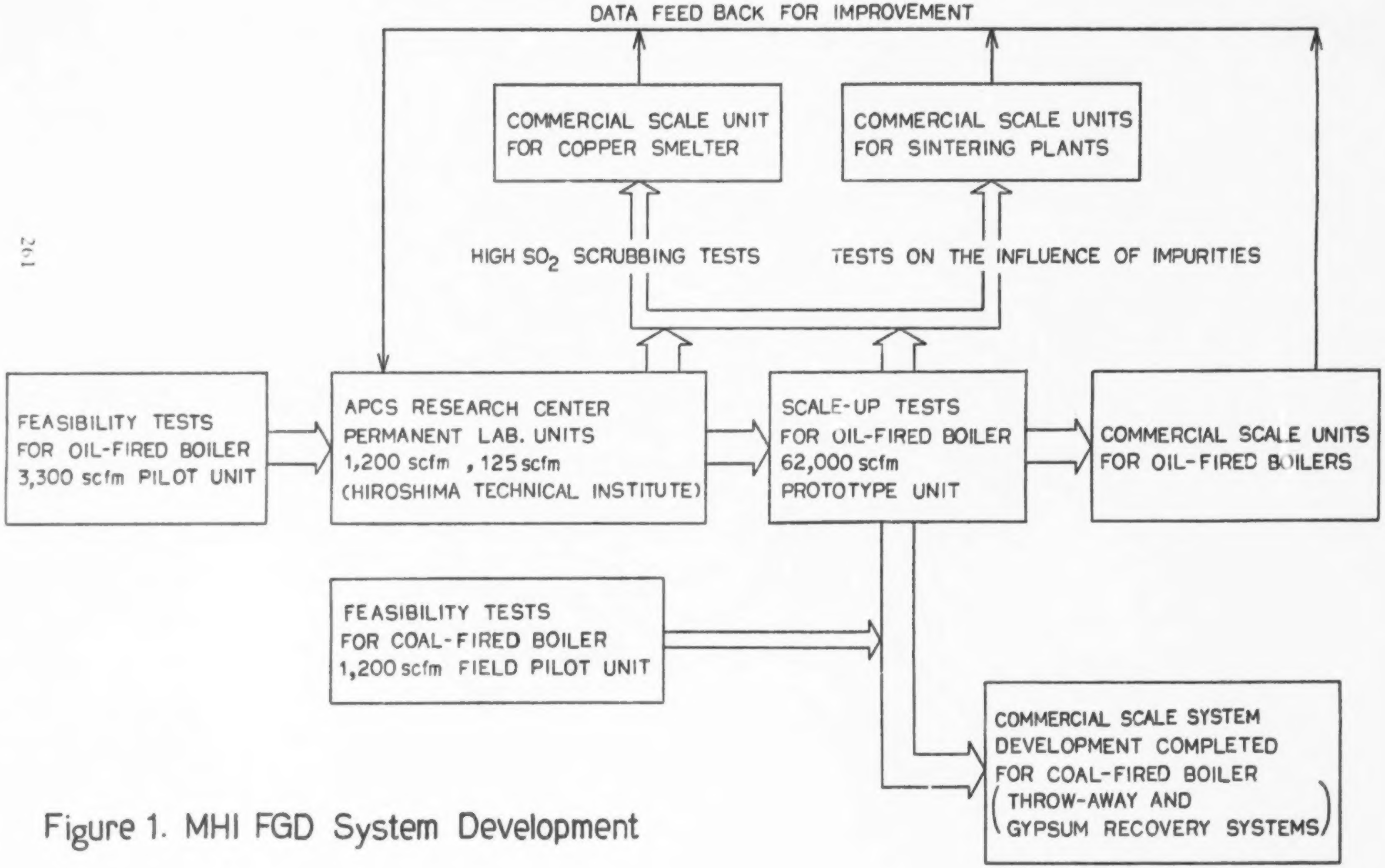

DATA FEED BACK FOR IMPROVEMENT
COMMERCIAL SCALE UNIT FOR COPPER SMELTER
COMMERCIAL SCALE UNITS FOR SINTERING PLANTS
HIGH SO2 SCRUBBING TESTS
TESTS ON THE INFLUENCE OF IMPURITIES
FEASIBILITY TESTS FOR OIL-FIRED BOILER 3,300 scfm PILOT UNIT
APCS RESEARCH CENTER PERMANENT LAB. UNITS 1,200 scfm , 125 scfm (HIROSHIMA TECHNICAL INSTITUTE)
SCALE-UP TESTS FOR OIL-FIRED BOILER 62,000 scfm PROTOTYPE UNIT
COMMERCIAL SCALE UNITS FOR OIL-FIRED BOILERS
FEASIBILITY TESTS FOR COAL-FIRED BOILER 1,200 scfm FIELD PILOT UNIT
COMMERCIAL SCALE SYSTEM DEVELOPMENT COMPLETED FOR COAL-FIRED BOILER (THROW-AWAY AND GYPSUM RECOVERY SYSTEMS)

Figure 1. MHI FGD System Development

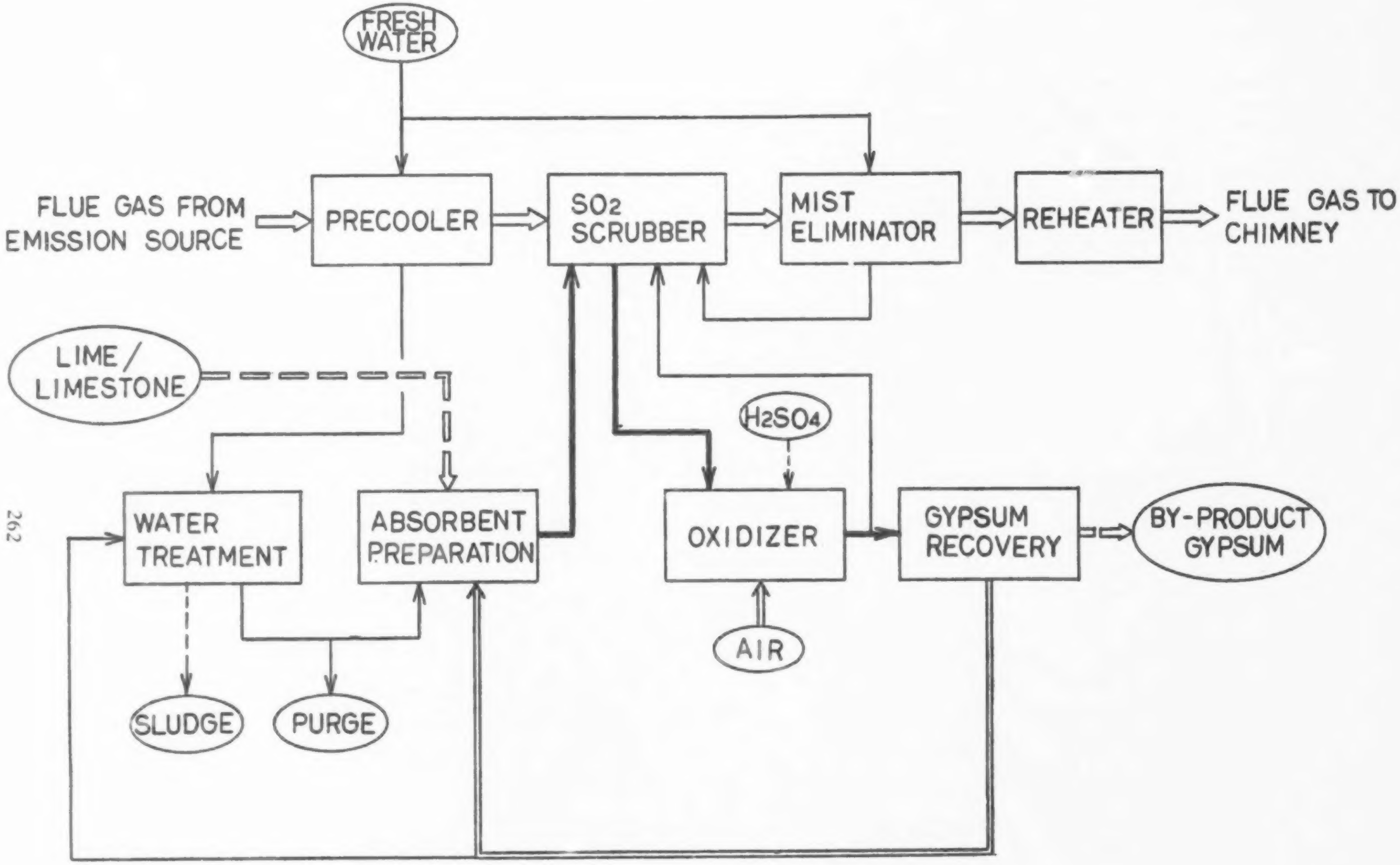

FRESH WATER
FLUE GAS FROM EMISSION SOURCE
PRECOOLER
SO_2 SCRUBBER
MIST ELIMINATOR
REHEATER
FLUE GAS TO CHIMNEY
LIME / LIMESTONE
H_2SO_4
WATER TREATMENT
ABSORBENT PREPARATION
OXIDIZER
GYPSUM RECOVERY
BY-PRODUCT GYPSUM
AIR
SLUDGE
PURGE

Figure 2. MHI FGD System (MHI Lime/Limestone-Gypsum Recovery Process)

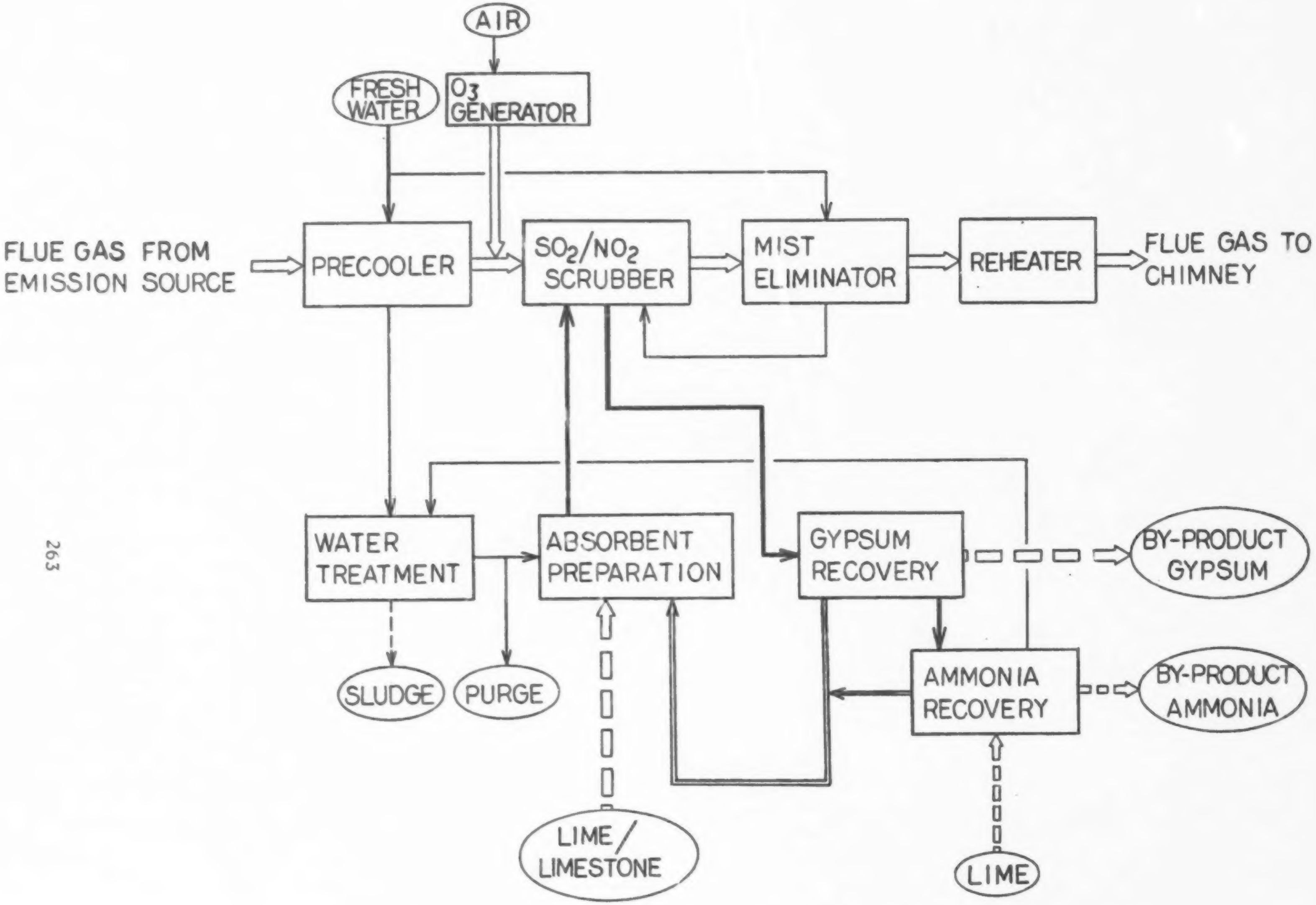

AIR
FRESH WATER
O_3 GENERATOR
FLUE GAS FROM EMISSION SOURCE
PRECOOLER
SO_2/NO_2 SCRUBBER
MIST ELIMINATOR
REHEATER
FLUE GAS TO CHIMNEY
WATER TREATMENT
ABSORBENT PREPARATION
GYPSUM RECOVERY
BY-PRODUCT GYPSUM
AMMONIA RECOVERY
BY-PRODUCT AMMONIA
SLUDGE
PURGE
LIME/ LIMESTONE
LIME

Figure 3. MHI Simultaneous NO_x/SO_x Removal Process

Table 1. DELIVERY RECORD OF MHI FGD SYSTEM

As of Jan., 1976

	Owner	Location	Gas Source	Capacity, scfm	Absorbent	Start-Up
1.	Kansai Electric Power Co.	Amagasaki	156 MW Unit	62,000	Lime	1972
2.	Onahama Refining Co.	Onahama	Copper Smelter	57,000	Lime	1972
3.	Yoshino Gypsum Co.	Tokyo	Industrial Boiler	8,000	Slaked Lime	1973
4.	Kawasaki Steel Corp.	Chiba	Sintering Plant	74,000	Lime	1973
5.	Kansai Electric Power Co.	Kainan	600 MW Unit	248,000	Lime	1974
6.	Tokyo Electric Power Co.	Yokosuka	265 MW Unit	248,000	Limestone	1974
7.	Tohoku Electric Power Co.	Hachinohe	250 MW Unit	235,000	Lime	1974
8.	Kyushu Electric Power Co.	Kanda	375 MW Unit	341,000	Lime	1974
9.	Kawasaki Steel Corp.	Mizushima	Sintering Plant	465,000	Lime	1974
10.	Kansai Electric Power Co.	Amagasaki	156 MW Unit	232,000	Lime	1975
11.	Teijin Limited	Ehime	Industrial Boiler	167,000	Lime	1975
12.	Kawasaki Steel Corp.	Mizushima	Sintering Plant	558,000	Lime	1975
13.	Kawasaki Steel Corp.	Chiba	Sintering Plant	260,000	Lime	1975
14.	Kuraray Co.	Saijo	Industrial Boiler	131,000	Lime	1975
15.	Mizushima Joint Thermal Power Co.	Mizushima	156 MW Unit	379,000	Lime	1975
16.	Toyobo Co.	Iwakuni	Industrial Boiler	124,000	Lime	1975

Table 1. (con'd)

As of Jan., 1976

	Owner	Location	Gas Source	Capacity, scfm	Absorbent	Start-Up
17.	Niigata Joint Thermal Power Co.	Niigata	350 MW Unit	328,000	Limestone	1975
18.	Chubu Electric Power Co.	Owase	375 MW Unit	744,000	Lime	1976
19.	Chubu Electric Power Co.	Owase	375 MW Unit	744,000	Lime	1976
20.	Kyushu Electric Power Co.	Karatsu	500 MW Unit	353,000	Limestone	1976
21.	Kyushu Electric Power Co.	Karatsu	375 MW Unit	452,000	Limestone	1976
22.	Kyushu Electric Power Co.	Ainoura	375 MW Unit	452,000	Limestone	1976
23.	Kyushu Electric Power Co.	Ainoura	500 MW Unit	452,000	Limestone	1976
24.	Tohoku Electric Power Co.	Niigata	600 MW Unit	260,000	Limestone	1976
25.	Kawasaki Steel Corp.	Chiba	Sintering Plant	446,000	Lime	1976
26.	Kashima-minami Joint Thermal Power Co.	Kashima	160 MW Unit	267,000	Limestone	1976
27.	Kansai Electric Power Co.	Amagasaki	156 MW Unit	294,000	Lime	1976
28.	Niigata Joint Thermal Power Co.	Niigata	350 MW Unit	328,000	Limestone	1977
29.	Sakata Joint Thermal Power Co.	Sakata	350 MW Unit	682,000	Limestone	1977
30.	Sakata Joint Thermal Power Co.	Sakata	350 MW Unit	682,000	Limestone	1977
31.	Kawasaki Steel Corp.	Mizushima	Sintering Plant	465,000	Lime	1977
32.	Chugoku Electric Power Co.	Shimonoseki	400 MW Unit	744,000	Limestone	1977

Table 2. OPERATING DATA OF MHI FGD SYSTEM

ITEM	EMMISION SOURCE	Oil-Fired Boilers	Sintering Plants (Steel Mills)	Copper Smelter
Inlet SO_2	ppm (Volume)	500 - 1,500	600 - 1,200	20,000 - 29,000
SO_2 Removal	%	90 - 99	90 - 97	90 +
Absorbent Utilization in Scrubber	%	90 - 99	95 - 99	- 99
Scrubber Control pH	-	6.5 - 7.0 (Lime) 4.0 - 5.8 (Limestone)	6.5 - 7.5	3 - 3.5 ① 9 - 10.5 ②
Oxidation in Scrubber	%	30 - 100	70 - 100	0 - 20
Slurry Concentration	% (Weight)	10 - 14	10 - 14	10 - 13
Inlet Gas Composition		(Pre-cooler Inlet)	(Pre-cooler Inlet)	(Scrubber Inlet)
O_2	% (Volume)	3 - 6	12 - 17	6
CO_2	% (Volume)	9 - 12	4 - 8	10
H_2O	% (Volume)	9 - 12	6 - 12	23
Dust	Grains/scf	0.0085 - 0.0175 (ESP Outlet) 0.04 - 0.08 (Boiler A/H Outlet)	0.0175 - 0.04 (ESP Outlet)	0.06
Remarks		One Scrubber System 12 Units Two Scrubber System 1 Unit	Inlet Cl : 20 - 50 ppm (as HCl) Inlet Oily Matter: - 0.025 gr/scf Dust Components: Fe, Mn, Si, Pb, K, Na, Ca, Mg, Al, Zn, Cu, etc.	Two Scrubber System ① No. 1 Scrubber ② No. 2 Scrubber

Table 3. AVAILABILITY OF MHI FGD SYSTEM

As of April, 1975

Owner (Location)	Start-Up (exclusive of test run)	Emission Source Operating Hrs (A)	FGD System Operating Hrs (B)	Availability (B)/(A) x 100 %	Remarks (Capacity: scfm x SO_2 ppm)
1. Kansai Electric Power Co. (Amagasaki)	Mar., 1972	16,000	14,600	91.5	Boiler stop : Apr. '74-Jan. '75 (62,000 x 800)
2. Onahama Refining Co. (Onahama)	Nov., 1972	24,120	23,780	98.6	(57,000 x 23,000)
3. Yoshino Gypsum Co. (Tokyo)	Aug., 1973	12,048	11,328	94.0	(8,000 x 1,500)
4. Kawasaki Steel Corp. (Chiba)	Nov., 1973	11,520	11,456	99.4	(74,000 x 500)
5. Kansai Electric Power Co. (Kainan)	Dec., 1973	8,760	8,760	100.0	Boiler stop : Dec. '74-Mar. '75 (248,000 x 300)
6. Tokyo Electric Power Co. (Yokosuka)	Apr., 1974	8,730	7,886	90.3	Boiler stop : May '74 (235,000 x 800)
7. Tohoku Electric Power Co. (Hachinohe)	May., 1974	4,800	4,080	85.0	Boiler stop : Dec. '74 (235,000 x 800)
8. Kyushu Electric Power Co. (Kanda)	Dec., 1974	2,880	2,880	100.0	(341,000 x 500)

References

1. Uno, T., M. Atsukawa, K. Muramatsu, 834 - 849, Proceedings of Second International Lime/Limestone Wet-Scrubbing Symposium, New Orleans, 1971.

2. Atsukawa, M., K. Matsumoto, N. Shinoda, M. Kimishima, Y. Ouchi, "Removal of SO_2 from Flue Gases by Wet Processes", M.H.I. Technical Review, February, 1974.

3. The data extracted from "Heavy Industries News", published in Japan, December, 1975.

4. Slack, A.V., "Scrubber survey: a lime/limestone trend", Electrical World, October 1, 1974.

Figures

1. MHI FGD System Development

2. MHI FGD System (MHI Lime/Limestone-Gypsum Recovery Process)

3. MHI Simultaneous NO_x/SO_x Removal Process

Tables

1. DELIVERY RECORD OF MHI FGD SYSTEM

2. OPERATING DATA OF MHI FGD SYSTEM

3. AVAILABILITY OF MHI FGD SYSTEM

STATUS OF FLUE GAS DESULFURIZATION USING ALKALINE FLY ASH FROM WESTERN COALS

Harvey M. Ness, Research Chemist
Everett A. Sondreal, Research Supervisor
Philip H. Tufte, Chemical Engineer

U.S. Energy Research and Development Administration
Grand Forks Energy Research Center
Grand Forks, North Dakota

ABSTRACT

It is the purpose of this paper to report on the status of flue gas desulfurization utilizing the alkali in fly ashes derived from Western coals. Most of the utility scrubbers operating in the Western United States have been installed for particulate removal. These particulate scrubbers, operating without an added reagent such as lime or limestone, have been found to remove appreciable amounts of sulfur dioxide from the flue gas and to experience sulfate scaling. Sulfur dioxide removal occurs because of the inherent alkalinity of the fly ashes produced from Western coals. The deliberate use of fly ash alkalinity for flue gas desulfurization has been investigated in four separate pilot-scale studies; two commercial-scale installations have been built and a third is being constructed. The properties of Western coals, their sulfur dioxide emissions, and the design and performance of selected commercial and pilot scrubbers are discussed in this paper. An appendix lists data on operating scrubbers.

STATUS OF FLUE GAS DESULFURIZATION USING ALKALINE FLY ASH FROM WESTERN COALS

INTRODUCTION

The Western reserve base for measured and indicated coal in place, as defined by the U.S. Bureau of Mines, totals 216 billion tons[1]. Future plans for mine expansion will increase capacity in the West above 200 million tons per annum by 1983[2]. This figure does not include tentative plans for a number of coal gasification plants. Considering these projections, stack gas cleaning technology for burning Western coals will assume much greater importance in the future than at present.

The Western coal reserves include lignite, subbituminous, and bituminous coal, with the lower rank coals predominating. An important property of almost all Western coals is that they contain far less sulfur than the 2 to 3 pct and higher typical of Eastern and Central coals. Unfortunately, sulfur content in Western coals (averaging 0.7 pct) is not generally low enough to meet new source emission standards. Typically, an average sulfur dioxide removal of 30 to 40 pct is required to meet the Federal standard of 1.2 lb SO_2/MM Btu, and higher removals are required to meet more stringent State and local standards. Since sulfur oxide emission standards are based on heat released, variations in heating value according to rank have an important effect on the coal sulfur content that is equivalent to the emission standard, as shown in Table 1.

TABLE 1 COAL SULFUR CONTENT EQUAL TO EMISSION STANDARDS

Coal	Higher heating value, Btu/lb	Coal sulfur equal to the Federal standard of 1.2 lb SO_2/MM Btu, pct
North Dakota lignite	6,800	0.41
Montana subbituminous	8,600	.52
Arizona (Black Mesa) bituminous	11,000	.66

While the 0.7 pct average sulfur content in Western coals does not satisfy the Federal standard, it does make stack gas cleaning potentially easier to achieve and less costly than for high-sulfur Eastern coals, provided that emission standards are not raised to cancel out the advantage. In addition, the alkaline nature of Western fly ash provides an opportunity for design innovations in flue gas desulfurization.

Ash content in Western coals can vary greatly, with the 4 to 20 pct shown in Table 2, representative of the overall range. The ash content varies significantly between mines, and even between locations within mines. The quantity of fly ash in stack gas depends on boiler design as well as on coal ash content. For a pulverized coal-fired boiler, fly ash leaving the boiler represents approximately 80 pct of the coal ash. Resulting particulate loadings are typically 2 to 10 gr/scfd. For a cyclone-fired boiler, fly ash leaving the boiler is approximately 40 pct of the coal ash. The corresponding particulate loadings are 1 to 5 gr/scfd.

A very important characteristic of many Western coal ashes is their high content of sodium oxide, magnesium oxide, and particularly calcium oxide (see Table 2). The alkali content tends to be highest in lignite, the lowest rank coal, and progressively less prevalent in the subbituminous and bituminous coals. Variations in alkali content are also influenced by the minerology of the overburden and the course of ground water movement. Alkali content in Western coal ash varies from under 10 to over 50 pct, with important variations occurring within individual mines.

A guideline in assessing the importance of the alkali in Western coal is the ratio of the alkali to coal sulfur. For a coal containing 7.5 pct ash and 20 pct alkali in the ash, the total alkali is chemically equivalent to slightly more than 120 pct of a 0.7 pct sulfur content. For some lignites, the alkali/sulfur ratio can be several hundred percent. Thus there is ample alkali to interact importantly with sulfur oxides in a wet scrubber in burning Western coals.

Required Removal Efficiencies

Required removal efficiencies are determined by coal sulfur content, sulfur retention on ash during combustion, and the emission limits established by law. As stated previously, the average 0.7 pct sulfur content of Western coals does not permit such coal to be burned without flue gas desulfurization under the Federal emission standard of 1.2 lb SO_2/MM Btu. Retention of sulfur oxides on ash during combustion may lower the sulfur dioxide emission by 10 to 40 pct for lignites[4], but this effect does not guarantee compliance with the Federal standard. Considering more stringent state and local standards, there are no coals that will meet the standards of Clark County, Nevada (0.15 lb/MM Btu), New Mexico (0.34 lb/MM Btu), Nevada (0.40 lb/MM Btu), or of Colorado after 1980 (0.35 lb/MM Btu); it is doubtful that any would reliably meet the Arizona statewide standards (0.8 lb/MM Btu). The removal efficiencies required to meet the more stringent standards are shown in Figure 1 for a Western subbituminous coal. At the 0.7 pct coal sulfur level, the required sulfur dioxide removal is increased from about 30 pct to meet the Federal standard to 90 pct to meet the Clark County, Nevada standard.

Capital and operating costs for stack gas cleaning must be expected to rise steeply as high percentage removals are required for substances that are initially present at low concentrations. For an approximate estimate based on principles of engineering design, it can be assumed that equipment size and power requirements for wet scrubbing will increase in proportion to the logarithm of one over the required exit concentration (size and power α log $\frac{1}{c}$ exit), and

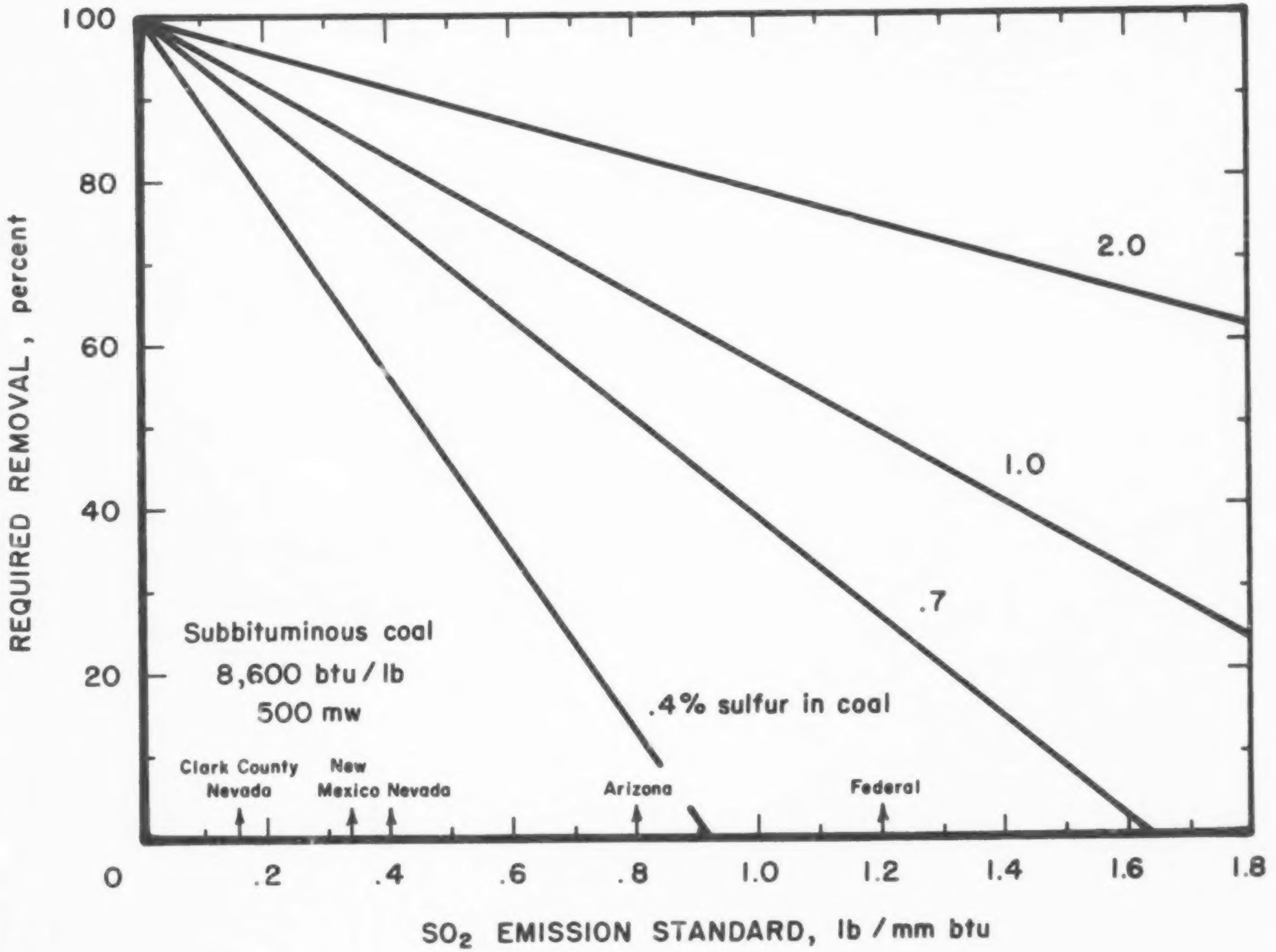


Figure 1. - Removal efficiencies required for stack gas cleaning of Western coal.

TABLE 2 SELECTED ANALYSES OF ASH IN WESTERN COALS[3]

Coal..........	Lignite	Subbituminous				Bituminous	
State.........	North Dakota	Montana	Wyoming	Arizona	New Mexico	New Mexico	Colorado
Mine..........			Big Horn	Black Mesa	Navajo	McKinley	Hawks Nest
Sample avg....	212	125	12	1	2	1	3
			Ash, percent of coal				
	6.2	9.3	4.8	7.5	20.2	8.0	5.4
			Oxide constituents, percent of ash				
SiO_2..........	19.7	35.5	27.4	42.0	55.6	54.7	44.8
Al_2O_3.........	11.1	18.7	12.7	18.1	26.2	21.6	28.3
Fe_2O_3.........	9.1	7.8	13.9	5.7	6.1	7.0	11.5
TiO_2..........	.4	.7	.6	.8	.6	1.0	.8
P_2O_5..........	.3	.3	.5	.6	.5	.0	.7
CaO...........	24.6	15.6	16.6	17.8	3.9	6.5	5.6
MgO...........	6.9	4.4	5.5	2.4	.8	1.2	1.9
Na_2O..........	6.5	1.7	2.2	1.4	1.5	1.6	.6
K_2O...........	.4	.4	.5	.3	.6	.8	.5
SO_3...........	19.5	13.4	17.0	8.2	3.2	5.8	4.0

capital cost will increase as the 0.6 power of size. Under these assumptions, equipment size would triple between 50 pct and 90 pct removal, and would double again at 99 pct removal. Capital cost would first double and then rise by a further 50 pct for the same increases in removal. These figures are hypothetical and they make no allowance for improved design or use of more effective reagents. However, they are close enough to reality to validly demonstrate that high costs will have to be paid to achieve improved control of stack gas emissions.

The following sections provide up-to-date information on the capital investment and operating experiences of the Western coal-burning utilites experiencing sulfur dioxide removal in particulate scrubbers. It also provides up-dated information on utility operated scrubbers that are deliberately utilizing fly ash alkali for sulfur dioxide removal, and also, on current research results on fly ash alkali utilization and scaling problems.

Arizona Public Service Company, Four Corners Plant[5, 6]

The Four Corners Plant at Farmington, NM has three pc-fired boilers (two 175 MW, and one 225 MW) that are equipped with Chemico venturi scrubbers for particulate removal, with two scrubber modules on each boiler. Two additional boilers of 755 MW each are equipped with electrostatic precipitators (ESP's). The first scrubber began operation in December 1971. Total cost of the scrubber system was $30 million, or $52 per kw.

The subbituminous coal burned at the Four Corners Plant, supplied by the Navajo mine, has as its outstanding characteristic, an unusually high ash content of nominally 22 pct. This results in a high dust loading in the flue gas leaving the boiler, 12 gr/scfd. Initial operation of the plant using mechanical collectors for fly ash removal resulted in dense plumes from the stacks, which dispersed to restrict visibility over a large surrounding region. Since scrubbers and ESP's have been operating, the plumes are greatly improved, although still visible.

The Navajo coal averages 0.68 pct sulfur, which produces a sulfur dioxide content of approximately 650 ppm. The fly ash alkalinity is low for a Western coal, having only 5 pct CaO in the ash. However, the amount of calcium entering the scrubber system is still relatively large, owing to the high percentage of ash. The calcium is chemically equivalent to approximately 75 pct of the coal sulfur content; and total alkalinity, including small amounts of Na_2O and MgO, exceeds 100 pct of sulfur equivalence.

A detailed description of the Four Corners scrubber installation is given in the Appendix and in Figure 2. Flue gas from the air heaters enters the venturi and then passes consecutively through a mist eliminator, a wet ID fan, another mist eliminator, and a steam reheater. However, because of erosion problems, the reheaters were removed and the units are operated as wet stacks. There is no bypass. Turndown is 50 pct.

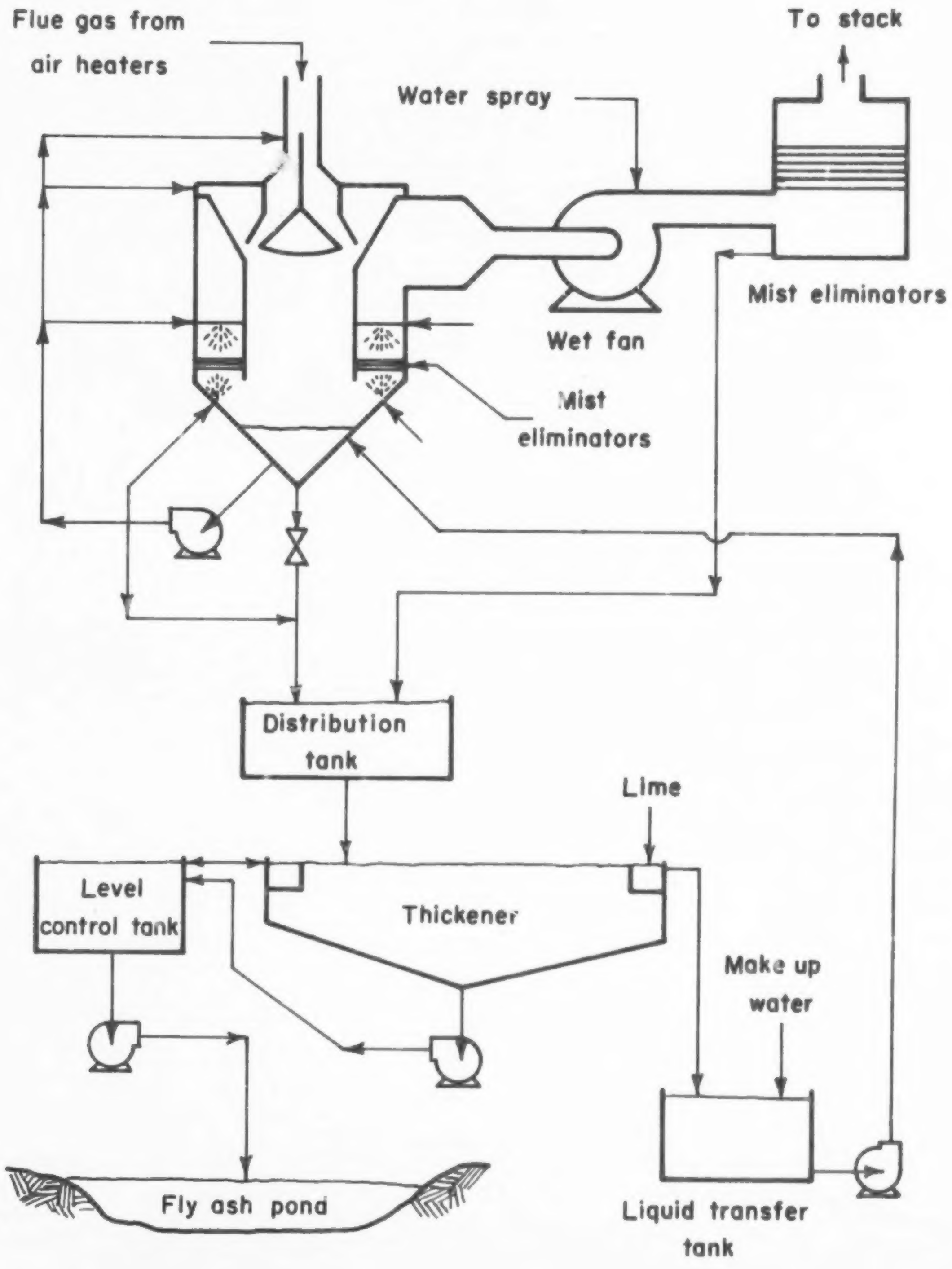

Flue gas from
air heaters
To stack
Water spray
Mist eliminators
Wet fan
Mist
eliminators
Distribution
tank
Lime
Level
control tank
Thickener
Make up
water
Fly ash pond
Liquid transfer
tank

Figure 2. - Simplified flow diagram for the Four Corners fly ash scrubbers.

Scrubber liquor is recycled back to the venturi and mist eliminator. The system was modified so that a bleed stream is taken from the recycle pump and sent to the thickener. Before the modification, blowdown was drained by gravity to the thickener. However, the gravity flow line was prone to plugging. Lime is now added to the overflow launders of the thickener instead of the centerwell, where some lime was lost by settling. The thickener and the fly ash transfer tank are at the same elevation and a line between them is used for water level control. The thickener underflow is pumped to a fly ash transfer tank and then pumped to a settling pond. The sludge is reported to have good settling properties. The pond is periodically dredged and disposed of in the mine.

The key operating variables are a liquid-to-gas ratio (L/G) of 18 gal/1000 scf, a pressure drop across the scrubber tower of 18-20 inches of water, and a pressure drop across the entire scrubbing system of about 28 inches of water. The venturi recycle liquid has a pH of 3.2 to 3.5. The pH of the thickener overflow liquid is maintained at about 7.5 by the addition of lime.

The operation of the venturi scrubbers has been satisfactory from the standpoint of meeting the particulate removal goal of 99.2 pct. Past measurements of sulfur dioxide removal, without supplemental lime, are reported to be in the range of 30 to 35 pct. The present lime addition rate is equivalent to about 7 pct of the sulfur dioxide in the inlet flue gas. Therefore, it would be expected to improve the sulfur dioxide removal by about 5 pct. The actual sulfur dioxide removal efficiency has not yet been measured. However, a long-range testing program with the primary purpose of reducing sulfur dioxide emissions, will be implemented by Arizona Public Service Company, since the existing level of sulfur dioxide emissions will not meet new sulfur dioxide limits. A secondary purpose of the test program will be to reduce scaling.

The testing program to be implemented will modify one scrubber system to test several variables. The variables to be tested are various liquid-to-gas ratios and suspended solids. Additional spray nozzles will be installed in the scrubber and an additional rubber-lined recycle pump will be added. The sulfur dioxide removal efficiency will be investigated as a function of L/G. The level of suspended solids will be maintained at 6 pct to provide seed crystals for precipitation of gypsum. Lime will be added for pH control.

Scrubber operating costs are not separated from the plant operating cost, and hence, are not available. The major operating requirements are electrical power equal to 3 to 4 pct of generating capacity, an estimated water usage of 5.9 acre-ft/MW/yr, manpower including 8 operators plus maintenance and supervision, and 14 tons of lime per day for control of pH.

Availability for the scrubber system is currently estimated at about 80 pct. This is the fraction of the time that a boiler is operating or could be operating that the scrubber modules are also operative. Since there is no bypass, this level of availability involves appreciable loss in power generation. With two scrubbers per boiler, lost generation can involve either reduction in load when one scrubber is inoperative or complete boiler shutdown when both scrubbers are inoperative.

The operating problems have been scale and solids buildup in the scrubber which would break off and clog the gravity-feed blowdown line. A severe erosion problem in the venturi opposite the plumb bell was resolved by installing silica carbide bricks. The bricks are reported to be very effective in resisting erosion. A problem of erosion on the tangential nozzles resulted in by-passing dirty flue gas to the wet fan. This, in turn, resulted in severe fan erosion. The problem was resolved by adding a stainless steel wear plate to the tangential nozzles. Additional problems are corrosion and leakage in the recirculatory lines and vessels and reheaters, where coatings on carbon steel has failed, on the mist eliminators and on the ducting leading into the stack. The blowdown line plugging has been resolved by using a forced bleed from the recycle pump. A manually-operated valve has been inserted in the old existing gravity blowdown line for emergencies. The reheater corrosion problem was resolved by removing the reheaters and using a wet stack. However, the stack liner and mortor were not acid resistant and, consequently, had to be replaced with the proper liner.

Scaling has occurred on most wetted surfaces, and it is not yet under control. Control measures include the use of an appreciable amount of blowdown (open-loop operation), an increase in the percentage of fly ash solids in the recirculating scrubbing liquor (from 2 to 6 pct), and addition of a new lime add system to maintain the pH at 7.5 in the thickener. It is not clear why this pH adjustment insures improved control of scaling, since in a system where the state of oxidation is high, with most dissolved sulfur present as sulfate, calcium sulfate would not be expected to be precipitated in the thickener by the rise in pH[7].

The measures being investigated for scale control at Four Corners, if successful, should find wide application in scrubbing applications involving low-sulfur Western coals.

GRAND FORKS ENERGY RESEARCH CENTER

The Energy Research and Development Administration's Grand Forks Energy Research Center (GFERC) has investigated the fly ash alkali sulfur dioxide scrubbing system since 1971. Testing has been performed on a 130 scfm pilot scrubber. The principal objectives have been to determine sulfur dioxide removal efficiencies and calcium sulfate scaling rates as a function of sulfur dioxide level, fly ash add rate, alkali in the fly ash, supplementary lime requirements, level of recirculated suspended solids, liquid to gas ratio, amount of makeup water and total dissolved solids. Past results have been published in three previous papers[8,9,10]. Research is continuing under funding from the Environmental Protection Agency, investigating the effect of high levels of total dissolved solids on sulfur dioxide removal efficiencies and calcium sulfate scaling.

The present GFERC scrubber (Figure 3) is a 130 scfm flooded disk venturi followed by an absorption tower containing conical "rain and drain" trays. Pressure drop across the scrubber can be controlled by adjusting the height of the flood disk. The conical trays were installed as a modification to increase the liquid to gas contact time. It was believed that this modificiation would eliminate the gas to liquid transfer step as a controlling variable at high

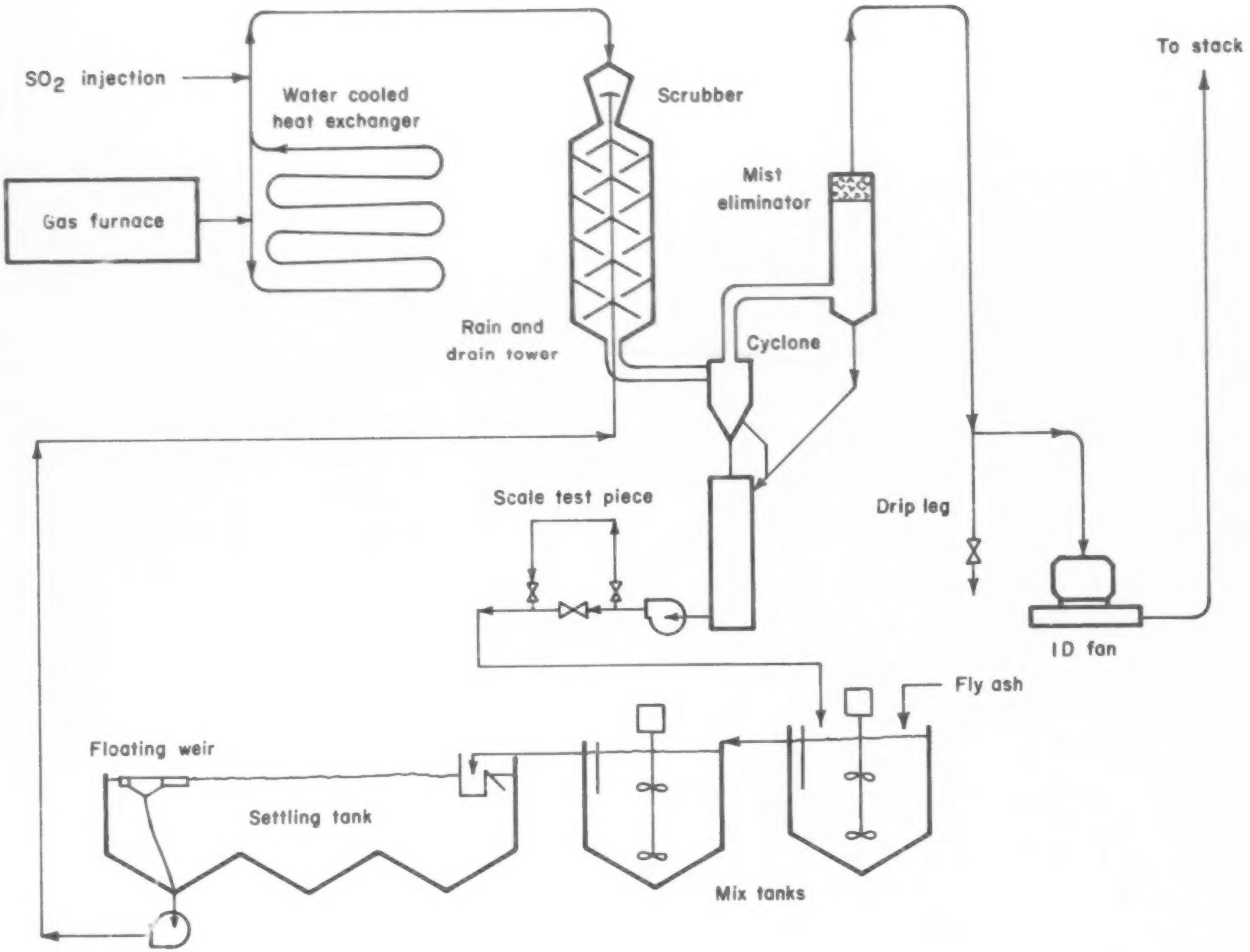

SO_2 injection
Water cooled heat exchanger
Gas furnace
Scrubber
Mist eliminator
Rain and drain tower
Cyclone
Scale test piece
Drip leg
To stack
ID fan
Fly ash
Floating weir
Settling tank
Mix tanks

Figure 3. - Pilot plant scrubber, Grand Forks Energy Research Center.

removal levels so that the sulfur dioxide removal observed would be a function of the solution characteristics only and not of scrubber design. Installation of the conical "rain and drain" trays increased the removal efficiency by 5 pct, from 83.3 to 88.1 pct under identical operating conditions.

The GFERC scrubber system is "closed loop." Recirculating scrubber liquor is lost from the system only as mist, which is equivalent to about 0.8 acre-ft/MW/yr, or as liquor in sludge. Efficient mist elimination is accomplished by passing gas through both a cyclone and a stainless steel wire mesh. Water lost by evaporation is replaced, but mist loss is not replaced during the course of a one-week test. The scrubber liquor is returned to a series of two fly ash mix tanks equipped with overflow weirs. The overflow from the second mix tank flows to a settling tank where calcium sulfate precipitates and unreacted fly ash is allowed to settle. A floating overflow weir in the settling tank provides the scrubbing liquid to the flooded disk.

Early experiments at GFERC indicated that a large increase in total dissolved solids, primarily sodium and magnesium sulfates, during the approach to steady state operating conditions significantly increased the scrubbing efficiency. Since the fly ash derived from some Western coals are known to contain significant amounts of soluble sodium and magnesium, it is probable that high concentrations of these species will result after long-term operation of a full-scale scrubber employing the fly ash alkali scrubbing process. Current experiments at GFERC are designed to investigate the properties of scrubber solutions that are high in sodium (0.5 to 10 pct) and magnesium (0.5 to 10 pct). The objectives of the current tests are to determine sulfur dioxide removal and scaling rate using a solution concentrated in sodium and magnesium and low in suspended solids (high levels of suspended solids, 6 to 12 pct, are common practice for scale control in many western scrubbers). The fly ash used in these tests contained high sodium and magnesium and was produced by pc-firing of Beulah, North Dakota lignite. Scrubber operating conditions kept constant for all test runs were: inlet sulfur dioxide level of about 840 ppm (a typical Western lignite containing about 0.8 pct sulfur), inlet flue gas temperature of 350° F, liquid temperature of about 120° F, absorber tower pressure drop of about 13 inches of water.

Scaling rates reported represent the rate of weight increase in grams per hour observed in a 3 ft 4 inch length of 1/2 inch I.D. pipe in the return line from the scrubber to the mix tanks. The test position chosen was a point of maximum scaling, and trends in the values observed were found to be well correlated with operating variables.

Tests were performed at liquid to gas ratios of 23, 45, and 75 gal/1000 scf. The CaO/SO_2 stoichiometric ratio was maintained at 1.2, sodium concentration at about 3.0 pct, magnesium concentration at 1 to 2 pct. The pH of the liquor pumped to the absorber tower varied from 5.0 to 5.5. Previous experiments indicated only a marginal effect when L/G was increased. However, under the conditions of high sodium and magnesium, removal efficiencies were affected very significantly. The removal efficiencies and fly ash alkali utilizations are tabulated in Table 3.

TABLE 3 SULFUR DIOXIDE REMOVAL EFFICIENCIES AND FLY ASH ALKALI UTILIZATION AS A FUNCTION OF L/G. CaO/SO_2 = 1:2

L/G	Removal Efficiency (pct)	Alkali Utilization (pct)
23	81.0	66
45	88.2	72
75	98.1	80

The stoichiometric ratios of calcium oxide to sulfur dioxide investigated were 0.6, 1.2 and 2.0. These ratios correspond to particulate loadings of 2.0 gr/scf, 4.0 gr/scf and 6.7 gr/scf. Operating conditions were as described above, with an L/G of 45. The sulfur dioxide removal efficiency, fly ash alkali utilization, and scaling rate are tabulated in Table 4. Removal efficiencies are shown in Figure 6.

TABLE 4 SULFUR DIOXIDE REMOVAL EFFICIENCIES, FLY ASH ALKALI UTILIZATION AND SCALING RATE AS A FUNCTION OF STOICHIOMETRIC RATIO, CaO/SO_2

CaO/SO_2	Removal Efficiency (pct)	Alkali Utilization (pct)	Scaling Rate (gm/hr)
0.6	63.3	100	1.68
1.2	88.2	72	2.7
2.0	98.0	49.1	3.0

In the above tests, some difficulty was experienced in removing the suspended solids to produce a "clear" liquid, even with the addition of sodium aluminate to the scrubber solution as a coagulant. The scaling rate of 1.68 gm/hr was observed at a suspended solids concentration of 0.13 pct, 2.7 gm/hr at 0.18 pct, and 3.0 gm/hr at 0.24 pct. Previous experience has shown that 10 gm/hr is a high scaling rate, and 0.2 gm/hr is a low scaling rate.

After the foregoing tests, the absorber tower was again modified, this time for the purpose of providing greater control of the pressure drop under conditions of severe scaling. The change involved attaching one set of cones, those directing flow from the center outward, to the standpipe of the flooded disk. Thereafter, movement of the standpipe varied the spacing between the convex and concave conical trays as well as the spacing of the flooded disk venturi. Thus, as scale buildup occurred on opposed surfaces, all such surfaces could be moved further apart to maintain a constant pressure drop. A further effect of the change was to distribute the pressure drop more evenly throughout the scrubber tower. This last effect was believed to be responsible for a further increase observed in scrubber efficiency, from 88.2 to 92.9 pct, which probably occurred because of a more efficient use of energy in redispersing droplets of scrubber liquor throughout the tower. All of the removal data given below are offset from former data due to this increased efficiency.

Scaling rates and removal efficiencies were next investigated as a function of sodium concentration. The sodium levels investigated were 0.17 pct, 0.66 pct, 4 pct, and 9.3 pct. The 0.17 pct represents sodium leached from the fly ash during a 3-day test period; no make-up sodium was added. The magnesium concentration was kept constant at 1 to 2 pct, L/G was 45, calcium oxide to sulfur dioxide stoichiometric ratio was 1.2, and other operating conditions were as described previously. The results are tabulated in Table 5. A typical solution analysis at each sodium level is listed in Table 6.

TABLE 5 SULFUR DIOXIDE REMOVAL AND SCALING RATE AS A FUNCTION OF SODIUM CONCENTRATION

Sodium Concentration (pct)	SO_2 Removal Efficiency (pct)	Scaling Rate (gm/hr)	Suspended Solids (pct)
0.17	95.2	5.8	0.066
0.66	92.9	3.51	0.074
4.0	93	2.7	0.17
9.3	96	.0	0.83

TABLE 6 TYPICAL SOLUTION ANALYSIS AT SODIUM LEVELS OF 0.17 PCT, 0.66 PCT, 4.0 PCT, AND 9.3 PCT

Species:	Percent Sodium:	0.17	0.66	4.0	9.3
Ca (ppm)		702	631	643	762
Mg (pct)		1.30	1.33	1.3	1.7
SO_4 (pct)		7.06	7.56	15.0	26.0

The solids settling properties were observed to degrade as the ionic strength of the scrubber solution increased. This phenomenon has been observed previously in EPA laboratory testing or dilute double alkali systems, and by Arthur D. Little, Inc.[11] in laboratory and pilot plant work on dilute and concentrated double alkali systems. Factors reported to influence the solids settling properties are reactor configuration, concentration of soluble magnesium and iron, and the concentration of soluble sulfate. In this investigation, the concentration of magnesium and iron were relatively constant. However, the level of soluble sulfate varied along with sodium level, due to the addition of sodium as sodium sulfate. The solids settling characteristics degraded correspondingly. A vacuum filter is being installed on the GFERC scrubber so that the level of suspended solids may be controlled more precisely.

It can be seen from Table 5 that the rate of scaling decreased as the sodium concentration increased. The absence of scale formations at the 9.3 pct sodium level is thought to be a function of sodium and not due to the higher (0.83 pct) level of suspended solids, since past work at similar levels of suspended solids (low sodium) resulted in scaling rates of 1 to 2 gm/hr. The stack flue gas was also tested to determine if sulfate was being lost in the mist. If sulfate was lost in the mist at a rate equal to or greater than that being absorbed into the scrubber solution (assuming constant liquid volume in the system), scaling would not be expected to occur. Extensive testing indicated this did not occur, and thus, the absence of scaling is concluded to be a result of the high sodium concentration.

It can also be seen from Table 5, that an increase in the level of total dissolved solids did not have a significant effect on sulfur dioxide removal, which contradicts previous results. The current result showing no effect on removal was observed after the modification of the scrubber by installation of "rain and drain" trays, which greatly increased gas-liquid contact in the scrubber. The conclusion to be drawn is that a high ionic strength in terms of sodium and magnesium sulfates increases removal for a scrubber configuration providing minimum contact-residence time (the flood disk venturi alone) but that it does not increase removal for a scrubber providing a maximum of contact-residence time (the venturi plus trays). The results would further indicate that the scrubber solution having low ionic strength had a sufficient equilibrium capacity to absorb essentially all of the entering sulfur dioxide (at 840 ppm and L/G = 45), but that this capacity was not fully utilized without the increased residence-contact time. On the other hand, the scrubber liquor having high ionic strength was indicated to have a greater affinity for rapidly absorbing sulfur dioxide such that essentially all entering sulfur dioxide could be removed with a short residence-contact time. Thus, the final conclusion is that sulfur dioxide removal in ash alkali scrubbing can be materially improved by either an increase in ionic strength or an increase in residence-contact time, but that a substantial increase in either can mask the effect of the other.

State of oxidation, of sulfite to sulfate, was generally high for all test conditions (98 to 99 pct sulfate). The percentage of sulfite was, however, higher in the test run at 9.34 pct sodium (2 pct sulfite) than in any other test.

Future tests will investigate the solution effects of varying the magnesium concentration at constant sodium levels and of scrubbing at low pH (below pH 3). In the present investigation, the level of magnesium was maintained at a constant value.

The GFERC is also working under a cooperative agreement with the Square Butte Electric Cooperative (SBEC), Minnesota Power and Light Company (MP&L), and Combustion Equipment Associates (CEA), to investigate the fly ash alkali scrubbing process using a 5,000 acfm (saturated) pilot scrubber. The objectives of the cooperative agreement are listed in the SBEC section.

Work is also planned on properties of scrubber waste from ash alkali scrubbing. The ideal properties for a scrubber waste are:

- absence of toxicity
- low soluble solids content
- low moisture content
- non-thixotropicity
- high compressive or bearing strength

Solid waste studied will be generated from operation of the 130-sfm scrubber and from SBEC's 5,000 acfm (saturated) pilot plant. Solubilities will be determined for the major soluble elements and for selected trace elements under conditions representative of disposal.

Minnesota Power & Light - Clay Boswell (Cohasset) and Aurora Stations[12]

Minnesota Power and Light operates Krebs-Elbair spray impingement scrubbers for particulate control on two 58 MW boilers at the Aurora Station and on one 350 MW boiler at Cohasset.

The Krebs-Elbair scrubber consists of a stainless steel box containing nozzles that direct a high pressure spray against baffles. The baffles consist of either vertical rods or a punch plate. The liquid is atomized when the spray impinges upon the baffles, and the resulting turbulance promotes an effective scrubbing of particulates. The Krebs-Elbair scrubber is designed to substitute power input from the high pressure spray for flue gas pressure differential, and it operates with only a 4-inch pressure drop across the scrubber system. This design supposedly affects a net operational savings in power and cost when compared to scrubbers employing relatively large pressure drops in the gas stream. An additional important feature is a nozzle tree design which permits sections of the nozzles to be removed for maintenance without scrubber shutdown. An apparent design limitation of MP&L's scrubbers is the erosion and plugging of the high pressure nozzles at other than very low levels of suspended solids. As previously stated, the scrubbers are designed primarily for particulate removal, although sulfur dioxide is removed because of the alkaline character of the fly ash particulate.

The coal burned at both the Aurora and Cohasset stations is a subbituminous from the Big Sky Mine, located near Colstrip, Montana. The coal typically contains 1.0 pct sulfur, 10 pct ash, and 9 to 13 pct CaO in the fly ash. The scrubber inlet particulate loadings are 2 gr/scfd at the Aurora station and 3 gr/scfd at the Cohasset station. A typical scrubber inlet sulfur dioxide level is about 800 ppm.

A detailed description of the MP&L scrubbers is given in the Appendix. A process diagram is shown in Figure 4. Flue gas from the air heaters passes through three concurrent sprays: a quench spray, high pressure spray, and the post humidification spray. The mist eliminator is between the high pressure spray and the post humidification spray. The post humidification spray washes the ID fan, which discharges flue gas to a wet stack. Scrubber liquor is pumped from a seal tank at the bottom of the spray chamber to two clarifiers. Overflow from the clarifiers is combined with makeup water and pumped back as spray. The ID fan is washed only with makeup water. Blowdown from the clarifiers is pumped to an eighty-acre ash pond. The scrubber operation is not closed loop. The Aurora station does not have clarifiers, and instead uses the clear liquor from the ash pond. In all other respects, the Aurora station is similar to that at Cohasset.

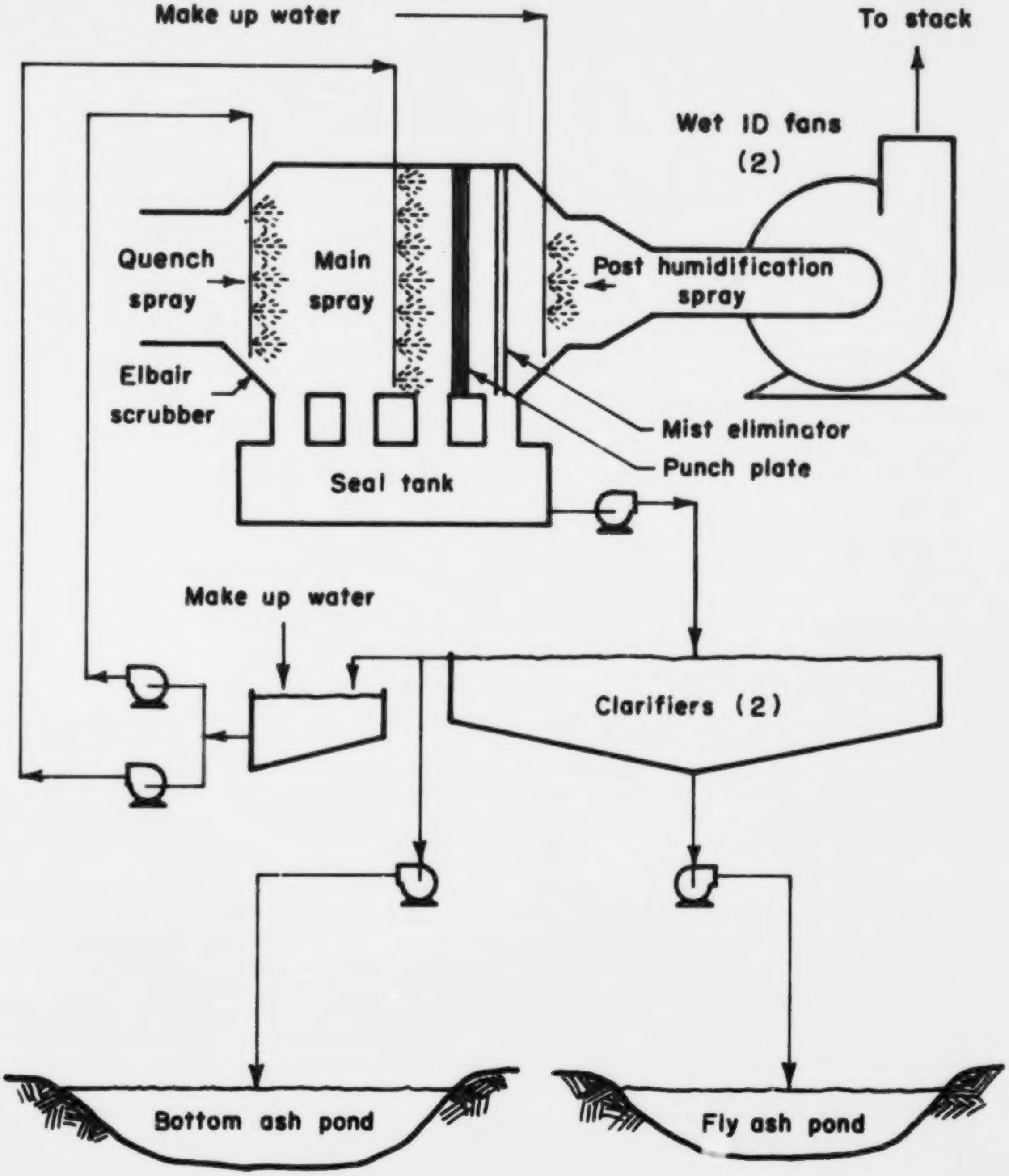

Make up water
To stack
Wet ID fans
(2)
Quench
spray
Main
spray
Post humidification
spray
Elbair
scrubber
Mist eliminator
Punch plate
Seal tank
Make up water
Clarifiers (2)
Bottom ash pond
Fly ash pond

Figure 4. - Simplified flow diagram for the Clay Boswell station particulate scrubber.

The key operating variables at both stations are L/G of 8.3 gal/acf and a total gas stream pressure drop of about 4 inches of water. The high pressure spray enters the scrubber at 200 psi. The pH of the liquid entering the scrubber is typically 4.4 and about 4.0 when leaving the scrubber. Particulate removal efficiency is about 98 pct at Aurora and about 97 pct at Cohasset. The sulfur dioxide removal is about 20 pct at Aurora and about 15 pct at Cohasset.

Operating costs are not available. Operating requirements are electrical power equal to 0.86 pct of generating capacity, water requirements of about 6.5 acre-ft/MW/yr at Cohasset and 30 acre-ft/MW/yr at Aurora, and a high labor requirement for maintenance and operation.

The scrubber availability was not obtainable in terms of effective scrubber operation during times that boilers were operational. However, little down-time on boilers would be expected using this scrubber process, since many scrubber problems can be repaired without boiler shutdown. Massive scaling, plugging, or problems involving the wet ID fan would, however, necessitate boiler shutdown.

The major problems that have occurred are stack gas mist carryover, scaling in the scrubber and liquid circuit, and heavy scale deposits on the wet ID fan and mist eliminators. The scaling problem was more severe at Cohasset, which operates close to its rated load and is restricted on the amount of blowdown. However, a seasonal variance had been obtained which allows greater blowdown, and the scaling and plugging problems have been less severe. About 420 gpm of clarifier underflow is pumped to the 80-acre fly ash pond. About 1000 gpm of sulfate-laden overflow from the clarifier is drained to the coal pile sump where it is diluted with drainage from selected plant drains, demineralizer waste water, boiler blowdown, stack drainage, and drainage from the coal pile and then pumped to the bottom ash pile pond and diluted with bottom ash slurry water. Some sulfate removal by precipitation is experienced. Overflow from the ash pond is pumped to the cooling water intake for units 1 and 2 and subsequently discharged into the Mississipi River.

The problem of stack gas mist carryover has been shown to be solvable by the use of a low velocity stack. A low velocity stack was simulated by breaching the brick liner of an existing high velocity stack. The flue gas could then flow through the annulus of the stack in addition to the brick liner and thus, in effect, simulate a low velocity stack. Subsequent tests on the modified stack indicated that mist carryover would not be a problem. An unexpected result of the low velocity stack tests was that the overall sulfur dioxide removal was increased from about 15 pct to 37 pct.

The increased sulfur dioxide removal in the wet stack was attributed to the wet ID fan and the modified low velocity stack. The water used to wash the ID fan is shattered into droplets which, depending on their size, would rapidly fall out or are transported to the stack. Depending on the transport velocity of the droplets, they could be carried out of the stack (which they were not), fall to the bottom of the stack, or be absorbed onto the walls, with some drops remaining in a dynamic suspension until becoming large enough to fall to the bottom of the stack. A situation has thus been created in which a water aerosal is continuously and intimately mixed with flue gas and fly ash aerosal. As a consequence, an additional 17 to 22 pct sulfur dioxide removal efficiency was obtained in the wet stack and the overall removal efficiency was increased to about 37 pct. The liquid that was drained from the stack to the bottom ash pond had a pH of about 3.

Minnesota Power and Light had planned on two 70 MW retrofit scrubber installations on units 1 and 2. However, MP&L now tentatively plans to use ESP's on the two 70 MW boilers and the Krebs-Elbair scrubber on the 350 MW unit and use a common low velocity stack. With all three boilers operating, the flue gas from the two 70 MW boilers would provide reheat to the 350 MW scrubber exit flue gas.

Montana Power Company Pilot Plant

A pilot plant program investigating ash alkali scrubbing was undertaken in 1973 for Montana Power by Bechtel Corporation in cooperation with Combustion Equipment Associates who designed the scrubber. Arthur D. Little, Inc. also participated in the test program, and sulfur dioxide and particulate testing services were provided by York Research Corp. Based on generally favorable results in the pilot plant program, Montana Power is employing ash alkali scrubbing in a new 362 MW pc-fired generating unit at Colstrip which started operation in September 1975. A similar unit is scheduled for startup in 1976 or 1977.

The information presented on these tests was obtained primarily from Dr. Carlton Grimm of Montana Power Company, and from a report published by Combustion Equipment Associates[13]. The present description of the tests as pieced together from the various sources is solely the responsibility of the writers of this paper.

The 3,000 acfm pilot plant (Figure 5) consisted of a venturi section for particulate removal, a spray tower for sulfur dioxide removal, followed by mist eliminators and a reheat section. Provisions were made for alkali makeup as Na_2CO_3 or lime and the use of cooling tower blowdown as makeup water.

The principal objectives of the study were to:

1. Demonstrate CEA's guarantees on sulfur dioxide and particulate removal.

2. Determine level of supplemental alkali required, if any, for sulfur dioxide removal.

3. Optimize variables influencing system performance.

4. Investigate use of cooling tower blowdown as a demister wash spray.

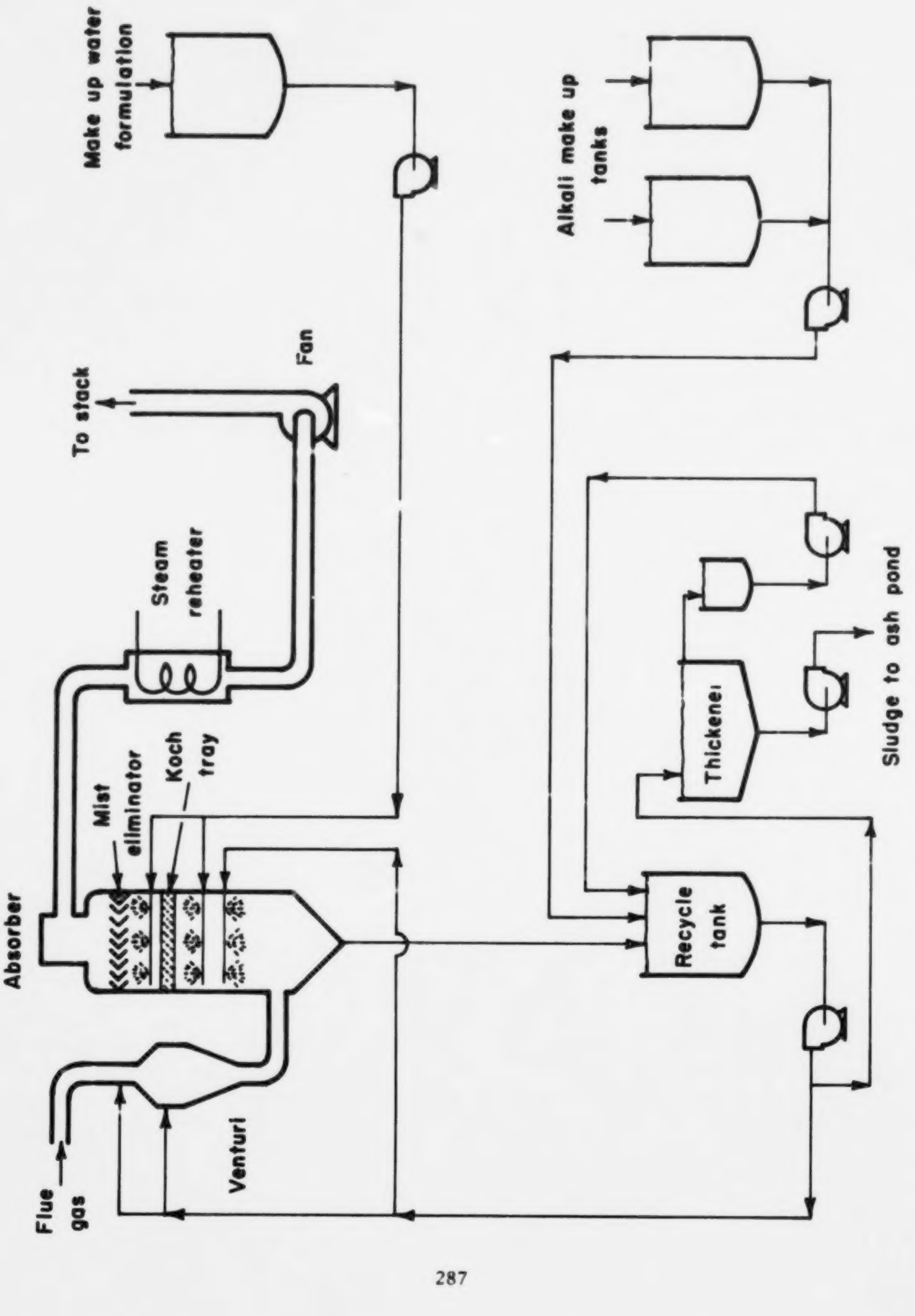

Flue gas
Venturi
Absorber
Mist eliminator
Koch tray
Steam reheater
Fan
To stack
Make up water formulation
Alkali make up tanks
Recycle tank
Thickener
Sludge to ash pond

Figure 5. - Pilot plant scrubber, Montana Power Company.

The goal for particulate removal was an outlet dust loading of 0.03 gr/scf. Results at an inlet dust loading of 2 gr/scf indicated that the 0.03 gr/scf exit loading could be met at a venturi pressure drops of approximately 12 inches of H_2O, and 0.02 gr/scf at about 17 inches of H_2O. A pressure drop of 17-inch H_2O was selected for the 4 to 6 gr/scf anticipated in the full scale units.

The sulfur dioxide removal requirement was to comply with a standard of 1 lb SO_2/MM Btu, or a level of 425 ppm. Compliance required removals in the range of 50 to 60 pct. This was accomplished with fly ash alone (without supplemental alkali) supplied at a dust loading of 2 gr/scf. At this dust loading, however, a pH of 4 in the scrubbing liquor entering the venturi was judged lower than desirable for scale control. Using a simulated grain loading of 4 gr/scf, a higher pH of 5 to 6 was achieved and the removal criterion was also met. An approximate plot of sulfur dioxide removal efficiency as a function of stoichiometric ratio alkali/SO_2 (inlet) is shown in Figure 6, for ash alone and with lime added.

A suspended solids level of 12 pct was thought to be adequate for sulfur dioxide removal and scale control, but higher levels would furnish more residence time for the ash to react and provide additional nuclei for precipitation.

Suspended solids level was found to have an important positive effect on sulfur dioxide removal in the range of 3 to 12 pct. Increasing L/G in the spray tower also increased sulfur dioxide removal. L/G and pressure drop in the venturi had little effect on sulfur dioxide removal.

Tests were run in which Na_2CO_3 was added for pH control. Sulfur dioxide removal was improved by 60 pct or more of the stoichiometric equivalent of the Na_2CO_3 added. There are no plans to use soda ash in the full scale units at Colstrip.

In other tests, lime was added to supplement the ash alkali. Sulfur dioxide removal was improved by 60 pct or more of the stoichiometric equivalent of lime added. (See Figure 6 for approximate results.) Use of lime is planned in full scale units to control pH during periods when fly ash loadings are less than 4 gr/scf or when fly ash reactivity decreases.

Scaling in the scrubber and liquid circuit was controlled adequately, mainly by the recirculation of ash. Some fouling did occur in the wet-dry zone, and loose scale deposit accumulated in the reheater.

Major emphasis was given to testing the chevron mist eliminators to establish their effectiveness and the wash conditions required to prevent plugging. Three approaches indicated that there was little measurable liquid entrainment through the demister. First, direct measurement of the degree of saturation of the flue gas indicated no carryover. Second, outlet particulate loadings with variable suspended and dissolved solids in the liquid showed no increase in outlet loading. Third, outlet particulate loadings did not change upon increasing or reducing demister wash sprays. However, deposits did form on the reheater, indicating that some carryover was occurring.

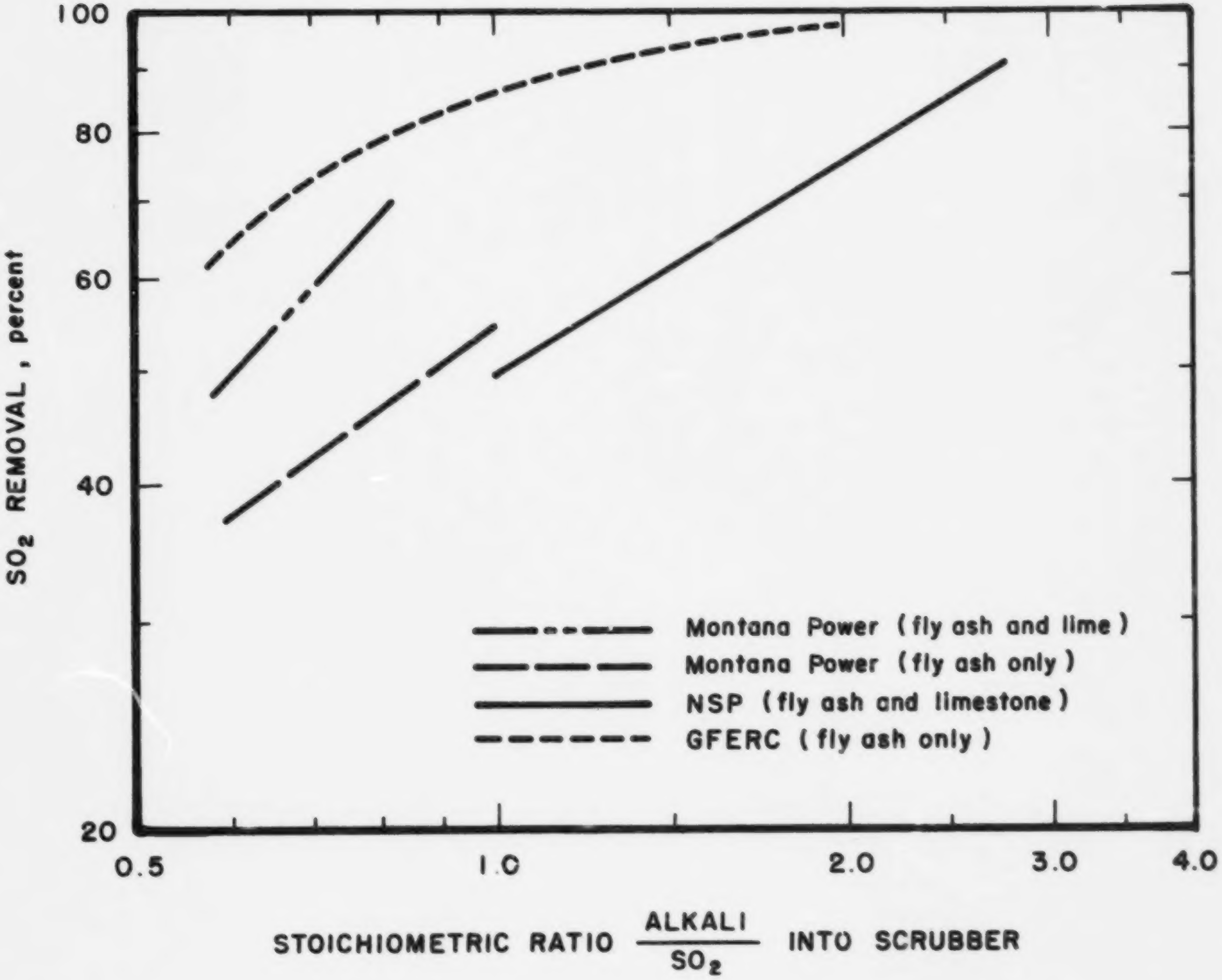


Figure 6. - Sulfur dioxide removals in pilot plant tests on alkaline fly ashes.

Simulated cooling tower blowdown was found to be unacceptable for washing the mist eliminators, because of scaling that occurred. To combat this scaling, a Koch bubbler tray was added ahead of the mist eliminator and reheater. Fresh makeup water used to wash the mist eliminator dropped into this tray, and re-entrainment of this relatively clean water from the Koch tray lowered suspended and dissolved solids in the mist by dilution and "exchange," aiding significantly in keeping the mist eliminator clean. Plans call for using this feature in the full scale units.

The level of dissolved solids in recycled scrub liquor was influenced by the amount and solubility of cations derived from the ash and by the amount and quality of makeup water added. Analysis at this end of a prolonged period of closed loop operations indicated that dissolved solids reached 26,000 ppm. Dissolved ion concentrations were as follows:

Magnesium	4,000
Chlorine	400
Sulfate	18,000
Sulfur trioxide	3,000
Calcium	400

The principal findings of the Montana tests were that acceptable sulfur dioxide removal could be achieved using fly ash alone as the source of alkali, and acceptable particulate removal could be obtained. Supplementary alkali could be used to control pH if fly ash quantity or quality were reduced. The optimum L/G for spray tower operation was determined to be in the range of 15 to 20 gal/1,000 scf.

Scale control was achieved principally through the recirculation of ash solids at 12 pct and above. Control of pH also aided in scale control. Other measures found helpful were fresh water washing of the mist eliminator and the use of the Koch bubbler tray. Cooling tower blowdown was found unacceptable as a source of makeup water for washing this mist eliminator. Operation of the full scale scrubbing unit at Colstrip has been considered successful to date.

Northern States Power Company - Black Dog Pilot Plant[14]

A study on fly ash alkali scrubbing was undertaken by Northern States Power, at their Black Dog Plant near Minneapolis, to establish the design of scrubbers for the Sherburne County Generating Plant where two 680 MW units are scheduled for completion in 1976 and 1977. Other parties involved in the test program included Combustion Engineering, the scrubber vendor, and Black and Veatch, consulting engineers.

The test facility was a marble bed scrubber with demister and reheat capability, having a capacity of 12,000 acfm. Attendant equipment included a reaction tank, thickener and ash pond, makeup and limestone tanks. Arrangement of equipment is shown in Figure 7.

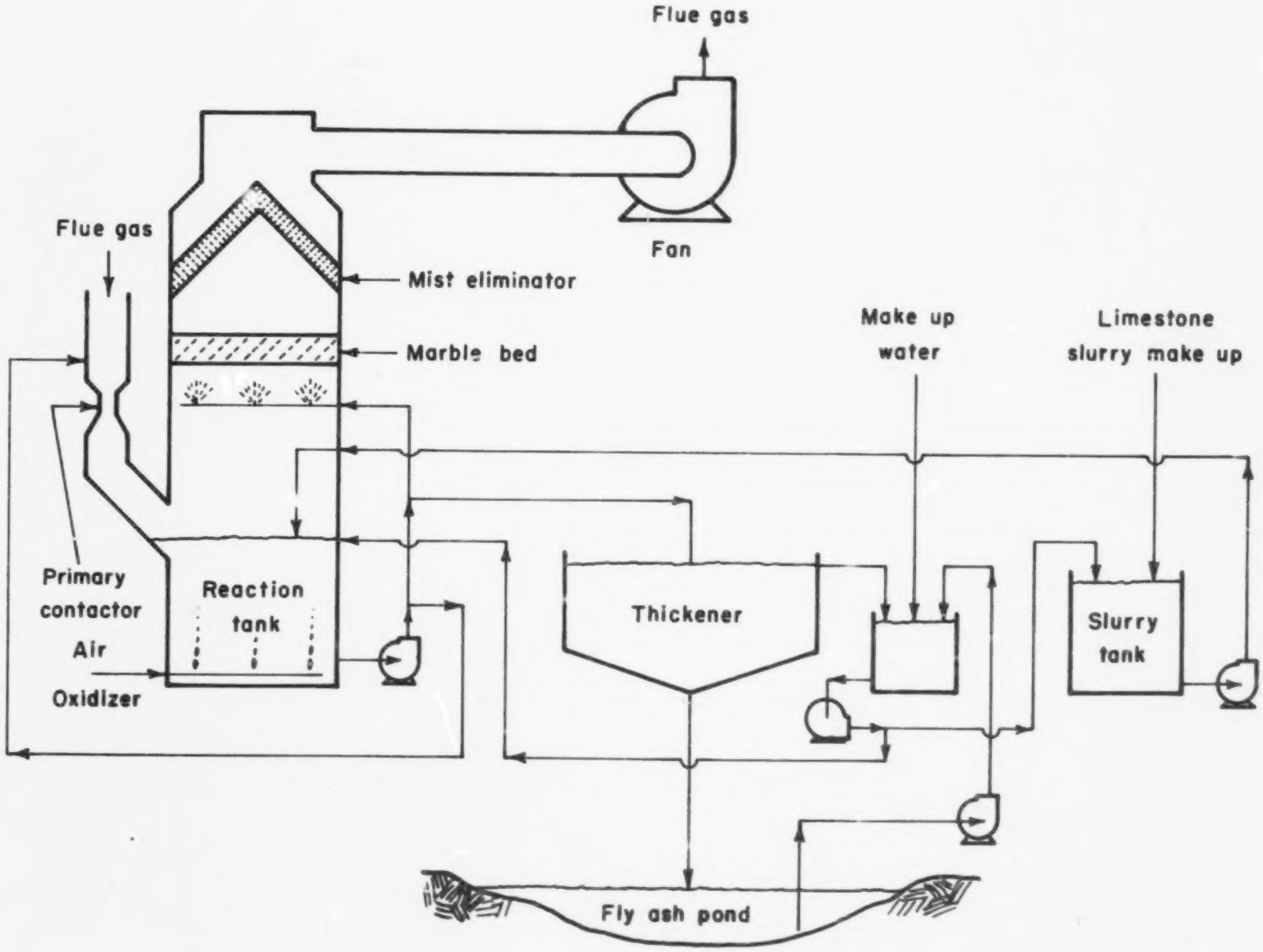

Flue gas
Fan
Flue gas
Mist eliminator
Marble bed
Make up water
Limestone slurry make up
Primary contactor
Reaction tank
Air
Oxidizer
Thickener
Slurry tank
Fly ash pond

Figure 7. - Pilot plant scrubber, Northern States Power.

The principal objectives of the test program were to:

1. Demonstrate capability for an SO_2 removal of 50 pct or greater operating on alkaline ash only or on ash and limestone.
2. Demonstrate the guaranteed particulate removal capability to 0.04 grams/scfd.
3. Demonstrate reliable operation with acceptable maintenance.
4. Determine optimum operating conditions, including blowdown requirements, and instrumentation and control functions.
5. Demonstrate reliable operation without deposit formation on the components.
6. Demonstrate acceptable continuous operation over a 30 day period.

To obtain the guaranteed dust removal of 0.04 gr/scfd outlet dust loading, or 99 pct removal, it was necessary to modify the marble bed scrubber by installing a venturi rod section (commonly called the primary contactor) at the inlet. The venturi rod section increased the total pressure drop from 6 to over 12 inches of water and the particulate criterion was met. Optimum system L/G was about 30 gal/1,000 acf. A demonstration test of the two-stage scrubber system for 50 continuous days was completed with 99 pct availability.

A series of tests were made wherein inlet sulfur dioxide and supplementary limestone were varied while the input fly ash remained relatively constant. The sulfur dioxide was varied from 482 ppm to 919 ppm and the limestone from 9 pct to 100 pct of stoichiometric. The sulfur dioxide removal efficiency varied from 48 to 88 pct under these conditions. The relationship between the stoichiometric ratio of input alkali to sulfur dioxide and the removal efficiency is shown in Figure 6. The sulfur dioxide removal criterion was met with as little as 5 pct stoichiometric limestone added. The calcium in the fly ash played a major part in the removal of sulfur dioxide, representing 70 to 80 pct of the alkali reagent.

Some scaling and plugging which occurred in the pilot program was remedied in the course of testing prior to the endurance test. Heavy mud deposits up to two inches thick appeared on the mist eliminator and scrubber wall surfaces, requiring shutdown every two days for washing. Deposits in both the mist eliminator and reheater were similar in chemical composition to the spray water solids, consisting primarily of fly ash with $CaSO_3$, $CaSO_4$, and $CaCO_3$. Deposits on the ID fan were chemically similar but physically more amorphous. Hard calcium sulfate scaling occurred on overflow pots. A redesign of the mist eliminator washing equipment was made to prevent plugging at that point. In addition, three modes of sulfate scale control were utilized. Ash solids were recirculated to provide ash and gypsum seed crystals, with a 10 pct level of ash found to be optimum. Supplementary limestone was reduced to a level of giving adequate sulfur dioxide removal while minimizing scale, with a level of approximately 15 pct of stoichiometric or less found to be optimum. A level of 30 pct stoichiometric was used in the endurance test. Oxidation of sulfite to sulfate was increased by bubbling air into the reaction tank to help prevent supersaturation with calcium sulfite in the scrubber proper. Oxidation level was increased from 36 pct to 98 pct and above with four times stoichiometric air. Good experience was reported with rubber-lined process equipment and fiberglass spray headers. slurry nozzles did not show wear or deterioration after 1700 hours of operation.

After 40 days of operation, the following concentrations of dissolved species in the spray water were measured:

	PPM
Sulfite	0
Sulfate	30,000
Chloride	550
Nitrate	400
Calcium	400-500
Magnesium	7,000
Silica	200
Sodium	200

Pacific Power and Light - Dave Johnston Plant[6, 15]

The Dave Johnston Plant at Glenrock, Wyoming has one 330 MW pc-fired boiler equipped with three parallel Chemico venturi scrubbers. The initial capital investment was $8 million, or $24/kw. A reported $5 million has been spent on improvements, bringing total cost to $39/kw. Startup was in April 1972.

Coal burned at the Dave Johnston Plant is Wyoming subbituminous coal from a captive mine. Sulfur content is 0.5 pct, resulting in a sulfur dioxide content of 500 ppm in the untreated flue gas. Coal ash content is 12 pct, and the calcium oxide content of the ash is approximately 20 pct. The calcium is chemically equivalent to 275 pct of the sulfur content. Inlet dust loading is 4 gr/scfd.

A detailed description of the Dave Johnston scrubber installation is given in the Appendix and Figure 8. Flue gas from the air heaters enters the venturi and then passes sequentially through mist eliminators, to a wet ID fan, and on to a wet stack. No reheat is used; there is no bypass. Turndown is to approximatel 30 pct of scrubber design capacity.

Scrubbing liquor is continuously recycled from the bottom of the venturi scrubber back to the plumb bob and to the deflector surrounding the bob that was installed to prevent solids buildup. Blowdown from this loop is pumped directly to two fly ash settling ponds; no thickener is used. Overflow from the settling pond is sent to a clear pond and then pumped back to the recycle loop. Ash is dredged from the settling ponds once each year and is hauled away for landfill.

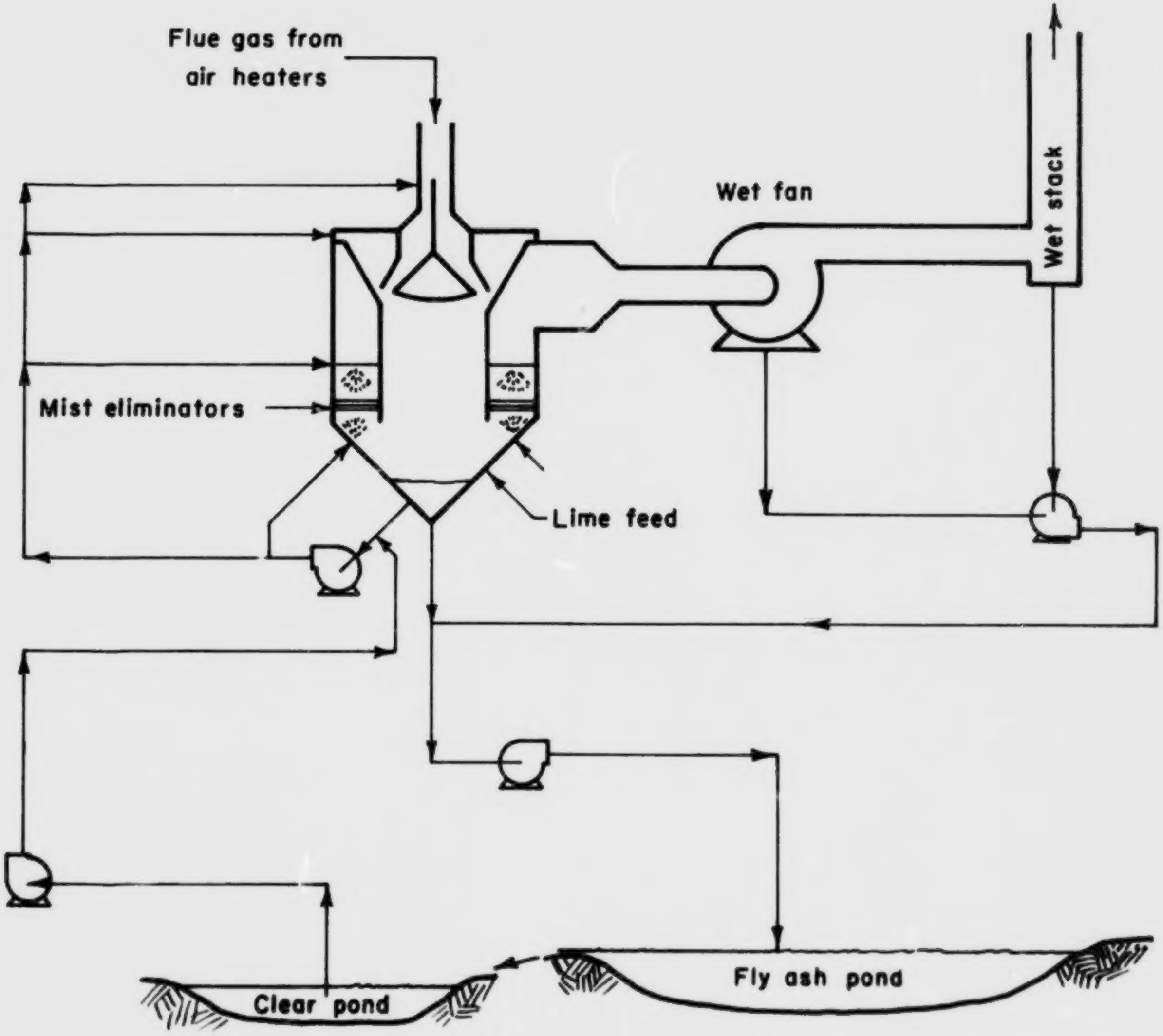

Flue gas from
air heaters
Wet fan
Wet stack
Mist eliminators
Lime feed
Clear pond
Fly ash pond

Figure 8. - Simplified flow diagram for the Dave Johnston fly ash scrubbers.

Key operating variables are an L/G of 13 gal/1,000 acf and a total pressure drop of 15 inches of water. The pH leaving the scrubber is 5, without supplemental lime.

Operating costs are not available. Major operating requirements are electrical power equal to 2.3 pct of generating capacity, an estimated water requirement of 3.6 acre-ft/MW/yr, lime for control of pH, and manpower.

Particulate removal efficiency is over 99 pct, meeting the design goal of 0.04 gr/scf. Preliminary values for sulfur dioxide removal are 40 pct without lime and somewhat higher with lime addition.

Availability is characterized by PP&L as being less than adequate for utility use, but no company-sanctioned percentages are available. Availability depends on the amount of blowdown and fresh water irrigation that are employed.

Operation is characterized as "intermittent open loop," meaning that operation with a minimum of blowdown is attempted as the normal mode of operation, with much larger amounts of blowdown and makeup used periodically to irrigate the system. Cooling water blowdown is mixed with service water (North Platte River water) and used for washing the ID wet fan.

Past major operational problem was scaling. At present, it is reported that scaling of the scrubber is still a major problem and an occasional fresh water irrigation is required. However, the previous detremental effects of wet-dry interface scaling have been minimized by the application of lignosulfonate, and is now not considered a major operational problem.

Future test plans are concerned with minimizing scale formation. The scrubber will be operated continuous with the recycle liquor pH maintained at 5.5 to 6.0, and scaling rates and cleaning requirements will be determined. The pH will be controlled using hydrated lime and feed rates are anticipated to be about 1,000 lb/hr. If the test is successful, a major development program will be implemented to design and install a permanent lime system utilizing pebble lime to maintain the pH at a constant level. The hydrated lime and future application of pebble lime will be to the scrubber vessel.

Public Service of Colorado - Valmont, Cherokee, and Arapahoe Stations[16]

The Public Service of Colorado has five turbulent contact absorbers (TCA) scrubbers containing thirteen modules installed on pc-fired boilers. The modules were designed by Universal Oil Products primarily for particulate removal. However, because of the alkali present in the fly ash, some sulfur dioxide is also removed. The modules consist of three stages of mobile packing, commonly called "ping pong balls," with the spray directed downward through the mobile packing, countercurrent to the gas flow. A booster fan directs the flue gas to the scrubber where it is presaturated and then passes through the mobile bed, then to chevron mist eliminators and on to a reheater. Reheat on most units is accomplished by heating the flue gas with steam coils; the Cherokee No. 4 unit uses externally heated air. All units have by-passes. Typical turndown capability is from 47 to 105 pct of the rated scrubber capacity. Detailed descriptions of the installations are given in the Appendix and Figure 9.

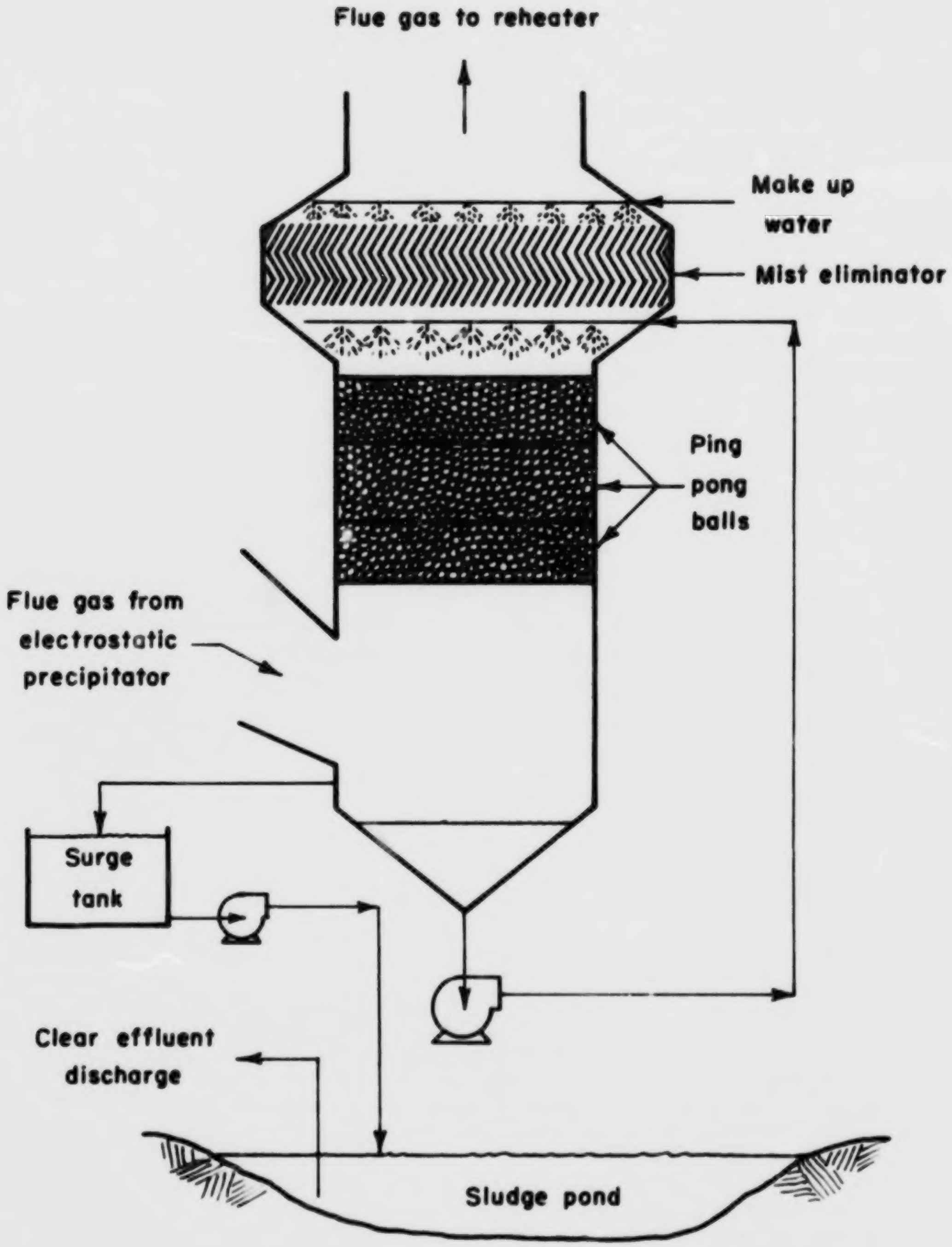

Flue gas to reheater
Make up water
Mist eliminator
Ping pong balls
Flue gas from electrostatic precipitator
Surge tank
Clear effluent discharge
Sludge pond

Figure 9. - Simplified particulate scrubbers at Valmont, Cherokee and Arapahoe Stations, Public Service of Colorado.

Public Service of Colorado's installation of the TCA scrubbers represents the last stage of particulate removal in a system which also uses mechanical collectors and electrostatic precipitators. All of the scrubber-equipped boilers are still serviced by the previously installed mechanical collectors and ESP's. At the Valmont Station, flue gas from the mechanical collector on a 196 MW boiler is split into two equal streams with one stream sent to a scrubber and the other stream sent to an ESP. The Cherokee Station, which has four boilers with a gross capacity of 115 MW, 170 MW, and 375 MW, has a particulate removal scheme in which the flue gas passes through, in series, mechanical collectors, ESP's, and then through scrubbers. The Arapahoe Station has a similar scheme.

The coal burned at the Valmont and Arapahoe Stations is Wyoming subbituminous and has a calcium content of 0.6 pct and an ash content of 5.2 pct. The calcium oxide content of the fly ash is about 20 pct, which is chemically equivalent to about 99 pct of the sulfur content. The Cherokee Station burns Colorado subbituminous coal which has a sulfur content of 0.7 pct and an ash content of 9.4 pct. The calcium content of the fly ash is about 5 pct, which is chemically equivalent to about 38 pct of the sulfur content. The dust loading at all the units range from 0.4 to 0.8 gr/scfd. The sulfur dioxide concentration at all the units is nominally 500 ppm.

Key operating variables at the Arapahoe and Cherokee Stations are a L/G of 54-58, a total pressure drop of 10 to 15 inches of water, an average pH of 7.0 entering the scrubbers and a pH of 2.8 to 3 leaving the scrubber. The scrubber liquor is recycled from the bottom of the scrubber to the spray header above the mobile bed. At the Arapahoe Station, slurry blowdown is pumped to an ash settling pond after adjusting the pH of the liquor to 6.0 to 9.0 with lime. The settled fly ash in the pond is dredged periodically and used for landfill. Clear effluent from the ponds is discharged under permit from the State of Colorado. At the Cherokee Station, slurry blowdown is neutralized and clarified and then mixed with the ash pond overflow for discharge. Scrubber operation at the Cherokee and Arapahoe Stations are considered open loop.

At the Valmont Station, a test program was implemented in October 1974 in which one of the two scrubbing modules was modified so that limestone for pH control could be added to the slurry feed. Each module receives 50 pct of the incoming gas, or 25 pct of the total gas flow. Previous operation of the scrubber system without pH control resulted in moderate to severe scaling, which required a high degree of maintenance and thus, poor availability. The scaling problem was a result, in part, of the high alkalinity of the Wyoming subbituminous coal which removed 45 to 50 pct of the inlet sulfur dioxide which, in turn creates supersaturated conditions with respect to calcium sulfate. Therefore, one module was modified so that limestone could be added to the slurry feed in order to adjust the pH to a range of 5.5 to 6.5. The level of suspended solids was desired to be about 7 pct to provide seed crystals for the precipitation of gypsum. Blowdown from the module is pumped to a lined settling pond. The settling characteristics of the sludge is reported to be good and contains about 50 pct moisture. Overflow from the pond is pumped to a recycle tank where makeup water is added at a rate of 70 gpm. The combined water is then used to slurry the limestone, wash demisters, and seal water for pumps. The only water loss in the system is due to pond evaporation, evaporation in the module, and the moisture in the sludge. The system is considered to be "closed loop."

Since the system has been operated for only four months, the results are considered to be preliminary in nature. Additional tests will be made in the future. The module operating without limestone removes about 45 to 50 pct of the incoming sulfur dioxide and has a particulate removal efficiency of about 96 pct. The modified module, using limestone for pH control, removes about 85 pct of the incoming sulfur dioxide and has a particulate removal efficiency of about 96 pct, which meets particulate emission requirements of 0.02 gr/scf (wet). The state of sulfur oxidation is high, reported to be nearly 100 pct. The degree of supersaturation is not known. Scaling in the modified module is reported to be moderate, although additional tests are required since some difficulty was experienced in maintaining the suspended solids at 7 pct The availability of the converted module was about 70 pct.

The particulate removal efficiency of the Cherokee and Arapahoe scrubber systems are 95 to 98 pct, which meets the design goal of 95 to 98 pct. The sulfur dioxide removal efficiency at the Cherokee Station, which burns Colorado coal, is 15 to 20 pct. Sulfur dioxide removal at the Arapahoe Station, which burns Wyoming coals, is 45 to 50 pct. The difference in sulfur dioxide removal efficiencies can be attributed, in part, to different fly ash alkalinity.

The scrubber operating problems experienced to date are wear and periodic replacement of the mobile bed packing, corrosive failure of the flue gas reheaters, and scaling. At Valmont, the modified scrubber module experienced moderate scaling on the module walls and certain portion of the first grid, which in turn caused pluggage of pump screens. Past experience at the other scrubber modules using additives added for scale control, have not been successful, although the tests were limited in nature. Scaling and plugging has occurred at the wet/dry zone on the first stage grids, and on the reheaters. With the exception of the Valmont experimental module, a higher than design blowdown is practiced in an effort to control scaling and prevent pluggage.

Direct operating costs for Cherokee #3 is 0.50 mill/kwh, based on 75 pct availability. Direct operating cost for the other stations are not available. Scrubber operating requirements are electrical demand equal to 4 pct of the power generated and also, steam for reheat. The water requirements are approximately 3.3 to 4.1 acre-ft/MW/yr. Manpower requirements for scrubber operations are not available. However, a high degree of maintenance is required.

Square Butte Electric Cooperative Pilot Plant

The Square Butte Electric Cooperative (SBEC) is currently constructing a 450 MW cyclone-fired generating unit requiring particulate and sulfur dioxide abatement controls. The 450 MW unit is referred to as Center 2 and is being constructed adjacent to the 238 MW Center 1 unit at the Milton R. Young Station. Particulate control will be provided by electrostatic precipitators (ESP's) and sulfur dioxide control by wet scrubbers. SBEC and Sanderson & Porter, Inc., consulting engineers, selected Combustion Equipment Associates (CEA) and Arthur D. Little, Inc. (ADL), to construct a 5,000 acfm (saturated) pilot plant and the full-scale scrubber based partly on previous pilot plant experience at Montana Power Company. The full-scale scrubber will utilize fly ash alkali with lime supplement. The design criteria and operating parameters of the full-scale scrubber were determined in two months of testing conducted by CEA-ADL in cooperation with SBEC, GFERC, and Minnesota Power and Light, on the 5,000 acfm pilot scrubber. The objectives of this program were:

1. To demonstrate that the stated performance guarantee could be achieved for a maximum coal sulfur content of 1.3 pct.
2. To determine design conditions for L/G, percent suspended solids in the recycle slurry, retention tank residence time, and pH of the recycle slurry.
3. To determine supplemental lime requirments at the maximum sulfur content.
4. To demonstrate the feasibility of using a flue gas bypass for reheat.

A long-range testing program using the 5,000 acfm (saturated) pilot plant will be conducted under a cooperative agreement between SBEC, Minnesota Power and Light Company, Combustion Equipment Associates, and the Grand Forks Energy Research Center. The objectives of the cooperative program are:

1. To determine whether sufficient alkali can be solubilized from cyclone-fired fly ash to reduce sulfur dioxide in flue gas below the level of State and Federal emission standards.
2. To determine the amount of additional alkali from lime which may be required to supplement fly ash alkali to meet State and Federal emission standards.
3. To determine the severity of calcium sulfite and calcium sulfate scale formation under normal operating conditions of the flue gas desulfurization pilot scrubber, and to investigate chemical methods of minimizing the scale formation.
4. To establish that the pilot scrubber can be operated on a closed-loop basis, and to determine the chemistry of the closed-loop system.
5. To determine what effect fly ash-derived soluble salts in the scrubber solution will have on the sulfur dioxide removal efficiency.

6. To determine and evaluate waste disposal of sulfate/sulfite sludge and fly ash-derived soluble salts in sludge.

7. To conduct corrosion tests to determine the effects of scrubber liquor on materials of construction to be used for full-scale flue gas desulfurization processes.

8. To determine the mass balance of all input and output materials, including selected trace elements and leachate from sludge.

9. To evaluate the capital and operating costs of fly ash alkali flue gas desulfurization for 100 MW, 500 MW, and 1000 MW steam generator plants based on the technical and operating data obtained from the pilot scrubber.

An additional phase of testing may be concerned with dilute sulfuric acid scrubbing with fly ash neutralization, and sodium-magnesium and calcium double-alkali-type scrubbing with fly ash neutralization.

A 5,000 acfm (saturated) pilot plant (about 1.4 MW equivalent) employing spray nozzles was chosen to minimize the pressure drop (hence, energy usage) across the absorption tower. The pilot plant (see Figure 10) has, essentially, two liquid loops: the primary sulfur dioxide scrubber loops, and the mist eliminator and wash tray loops.

The wash tray loop operates on clear liquor at approximately pH 4 and consists of a wash tray above the absorber tower, a tray recycle tank, clarifier and clarifier overflow tanks, and a demister. The wash tray, which is constructed of 316L stainless steel, is designed to remove entrained droplets of scrubbing liquor which could otherwise foul the demister. Liquid from the wash tray drains to an 8 x 8-foot flakeglass-lined recycle tank which is then pumped back to the wash tray or to an 8 x 8-foot flakeglass-lined clarifier. Overflow from the clarifier is used to wash the bottom of the wash tray. Liquid from the clarifier not used for washing is drained by gravity to a 6 x 6-foot flakeglass-lined overflow tank. Makeup water from nearby Lake Nelson is added to the pilot scrubber at the clarifier overflow tank at an average rate of 1.4 gpm (about 1.6 acre-ft/MW/yr) and the combined liquid is used to wash the polypropylene demister. Underflow from the clarifier is pumped to a 316L stainless steel vacuum filter.

The scrubber loop operates on a slurry of alkali ash in recycled liquor at a characteristic pH of 4 to 7. It consists of a 45 foot high by 3 1/2 foot diameter flakeglass-lined absorber tower which contains six banks of 316L stainless steel nozzles spraying scrubber liquid countercurrent to the gas flow. The scrubber liquid drains to a 12 x 8-foot flakeglass-lined retention tank equipped with a 316L stainless steel agitator. The retention tank liquid is pumped back to the spray nozzles and also to an 8 x 8-foot flakeglass-lined thickener which is used to control the level of suspended solids. Reducing the liquid flow from the retention tank to the thickener will increase the level of suspended solids;

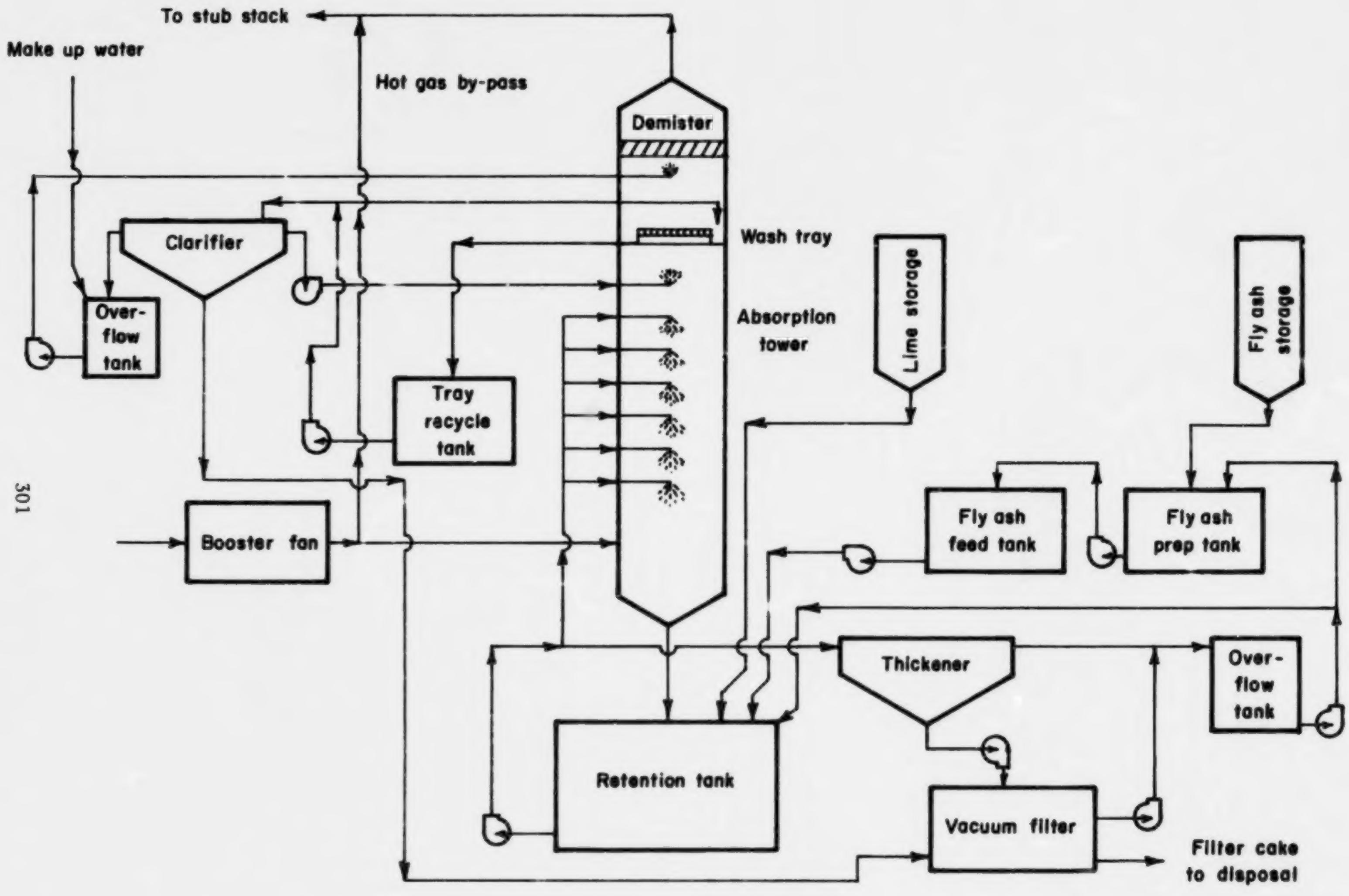

To stub stack
Make up water
Hot gas by-pass
Demister
Clarifier
Wash tray
Over-flow tank
Absorption tower
Lime storage
Fly ash storage
Tray recycle tank
Booster fan
Fly ash feed tank
Fly ash prep tank
Thickener
Over-flow tank
Retention tank
Vacuum filter
Filter cake to disposal

Figure 10.- Flow diagram of 5,000 acfm saturated pilot plant scrubber, Square Butte Electric Cooperative.

increasing the flow rate will lower the level of suspended solids. Overflow from the thickener drains by gravity to a 5 x 5-foot flakeglass-lined overflow tank where it is mixed with liquid pumped from the vacuum filter. Thickener underflow is pumped to the vacuum filter. The vacuum filter is operated only when the concentration of suspended solids in the thickener underflow is above approximately 40 pct. The liquid from the thickener overflow tank is pumped to a 4 x 5-foot fly ash preparation tank with the excess liquid returning to the retention tank. The fly ash slurry is pumped to an 8 x 8-foot feed tank for additional mixing time, and then to the retention tank. Fly ash is stored in a 3 x 5-foot hopper and fed to the preparation tank using a screw feeder at rates up to 8 lb/min. Lime from a 3 x 2-foot storage hopper and screw feeder is used for pH control and is fed directly into the retention tank without slaking. The total amount of liquid in the entire pilot plant scrubber is about 25000 gallons. All pumps are rubber lined. Liquid flows are measured by rotometer; liquid and gas temperatures are measured by dial thermometers, and pressure drops are measured by manometers.

The pilot plant scrubber has the necessary equipment and controls to operate over a wide range of variables. The solution pH can be varied from below pH 4 up to pH 9; the retention tank residence time may be varied from 4 minutes to 16 minutes; the liquid to gas ratio may be varied from 40 to 80; and a sulfur dioxide injection system can adjust the scrubber inlet sulfur dioxide to any desired concentration. A duct equipped with an orifice and damper has been installed between the inlet and outlet of the absorption tower and can be used to bypass part of the hot inlet flue gas to mix with the cooler outlet flue gas leaving the absorption tower. The mixing of flue gases in this manner is being tested as a method for reheating to a temperature above the saturation point to eliminate the possibility of stack gas rain.

A mobile trailer, constructed by KVB Industries according to specifications supplied by the Grand Forks Energy Research Center, provided the capability of continuously monitoring both the inlet and outlet flue gas for sulfur dioxide, nitrogen oxides, carbon dioxide and oxygen. In addition to the gas monitoring equipment, the trailer contains a chemistry laboratory to perform most analyses of coal and scrubber liquor on site.

The preliminary results of the test program completed to date have been favorable. The sulfur dioxide removal is required to comply with the Federal emission standard of 1.2 lb/MM Btu, which corresponds to approximately 535 ppm SO_2 (dry). The sulfur dioxide removal efficiency was investigated as a function of L/G, suspended solids, inlet sulfur, sulfur dioxide concentration, and fly ash add rates. The ESP inlet fly ash particulate loading at the inlet to the ESP on Center 1 ranges from 0.71 to 1.53 gr/scf and averages 1.13 gr/scf. The three ash add rates investigated were equivalent to the average amount collected by the ESP on Unit 2, the combined average amount collected by the ESP's on Units 1 and 2, and the maximum amount collected on Units 1 and 2. A typical analysis of the fly ash is shown in Table 7.

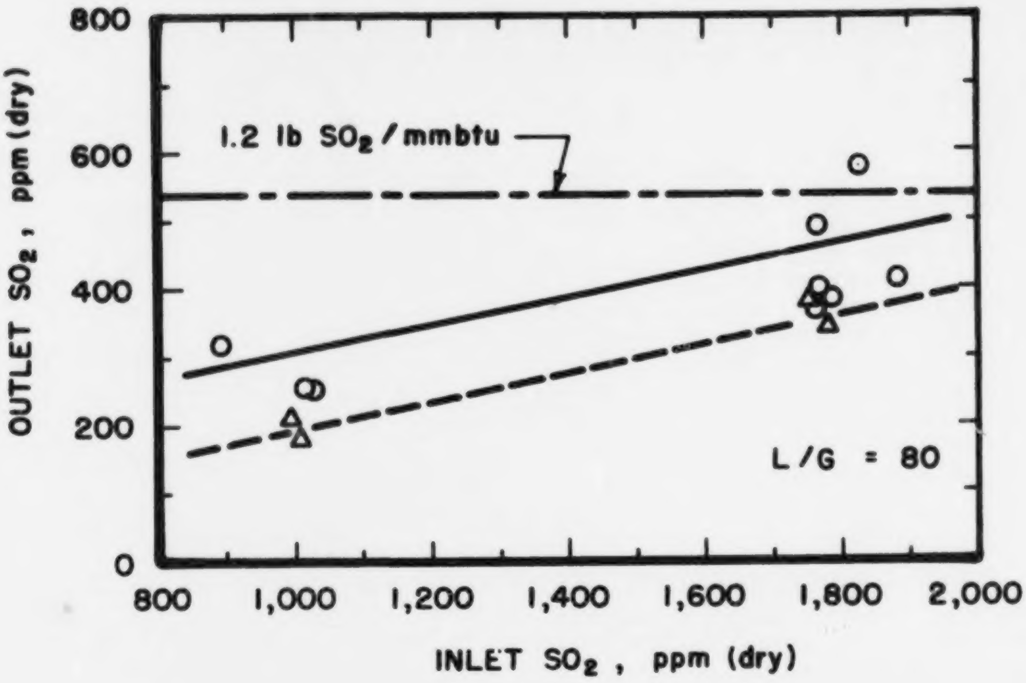
TABLE 7 A TYPICAL ANALYSIS OF THE FLY ASH PRODUCED BY THE CYCLONE-FIRED CENTER UNIT NO. 1

	Percent of ash, as received
Loss on ignition at 800° C	2.2
Silica, SiO_2	29.8
Aluminum oxide, Al_2O_3	12.7
Ferris oxide, Fe_2O_3	10.6
Titanium oxide, TiO_2	0.5
Phosphorous pentoxide, P_2O_5	0.3
Calcium oxide, CaO	25.7
Magnesium oxide, MgO	4.5
Sodium oxide, Na_2O	2.2
Potassium oxide, K_2O	2.0
Sulfur trioxide, SO_3	6.4
TOTAL	96.9

Figure 11 illustrates the sulfur dioxide removal efficiency at the above fly ash add rates at L/G ratios of 60 and 80. The dashed line corresponds to the average fly ash production collected by both units, hereafter referred to as the average ash add rate. The solid line corresponds to the maximum fly ash production by both units, hereafter referred to as the maximum ash add rate. A sulfur dioxide level of about 1100 ppm (dry) would be equivalent to about a 0.75 pct sulfur coal (HHV6604). A level of 1900 ppm (dry) is equivalent to about 1.3 pct sulfur in coal. The oulet sulfur dioxide represents the removal for the total scrubber system, which includes the flue gas by-passed and used for reheat. The total flue gas into the system was 6300 acfm, of which 1100 acfm by-passed the absorber tower to provide reheat. The inlet gas temperature was about 325° F. The temperature of the saturated gas out of the absorber tower was about 135° F. After mixing the by-pass gas, the temperature of the gas to the stack was about 155° F. The flue gas reheat was tested as an alternative to coil reheaters. No stack gas mist was observed to occur.

At a L/G of 60, an inlet level of about 1100 ppm SO_2 (dry) using fly ash at the average add rate, the sulfur dioxide removal efficiency for the total scrubber system was about 61 pct (absorber tower was about 70 pct). The total fly ash alkali utilization, based on 25 pct CaO, 4.72 pct MgO, and 1.67 pct Na_2O, is about 63 pct. At the maximum ash add rate, the sulfur dioxide removal is about 69 pct (absorber tower was about 84 pct), which represents a total alkali utilization of 51.5 pct. Supplemental lime was not added and the corresponding pH of the recycle slurry was about 3.9 at the average ash add rate and 4.9 at the maximum ash add rate.

At a L/G of 60, inlet level of about 1850 ppm SO_2 (dry), using the average fly ash add rate with lime supplement, the total scrubber system removal efficiency was about 70 pct (absorber tower removal of about 84 pct). The supplemental lime was added to maintain the pH at 6.0 to 7.0, and was chemically equivalent from about 90 to 100 pct of the 1850 ppm inlet sulfur dioxide; the total alkali



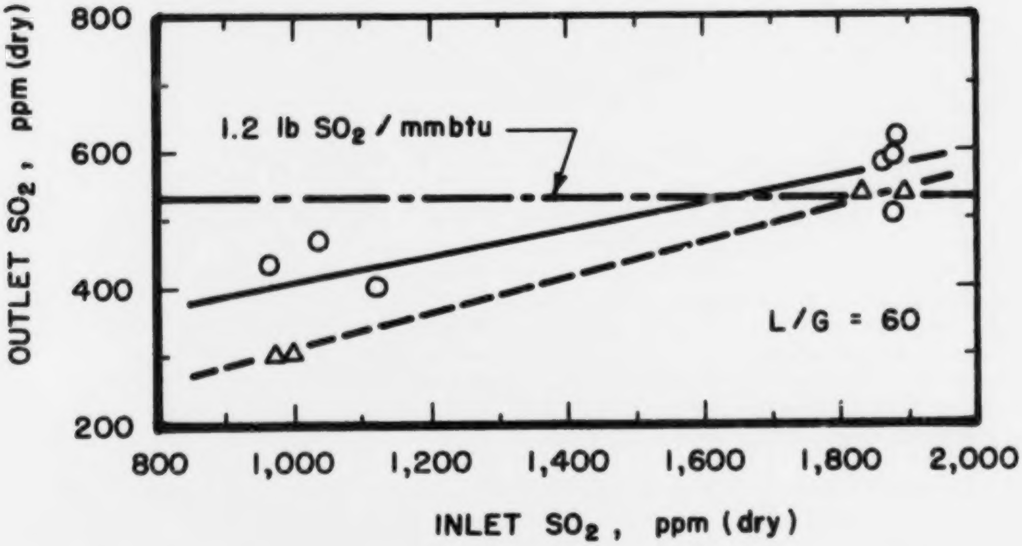


Figure 11. - Sulfur dioxide removals in SBEC pilot plant tests using fly ash alkali. Solid line represents average fly ash (1.13 gr/scf) collected by ESP'S and dashed line represents maximum fly ash (1.53 gr/scf) collected by ESP'S.

(fly ash alkali and lime supplement) was equivalent to about 140 pct of the inlet sulfur dioxide. At the maximum ash add rate with lime supplement, the sulfur dioxide removal remained at about 70 pct (absorber tower about 84 pct). The supplemental lime was added to maintain the pH at 6.0 to 7.0 and was chemically equivalent to about 68 pct of the inlet sulfur dioxide; the total alkali (fly ash alkali and lime supplement) was equivalent to about 142 pct of the inlet sulfur dioxide. In a separate test using only lime chemically equivalent to that of the total alkali, the removal efficiency increased to about 79 pct.

At an L/G of 80, inlet level of about 1100 ppm SO_2 (dry) using fly ash at the average add rate, the sulfur dioxide removal efficiency for the total scrubber system increased to about 84 pct (about 86 pct in the absorber tower). The total fly ash alkali availability was increased to about 70 pct. The pH of the recycle slurry was about 3.8. No supplemental lime was used. At the maximum ash add rate with supplemental lime, the total scrubber system sulfur dioxide removal was about 81 pct (absorber tower was about 97 pct). The supplemental lime was added to maintain the pH at 6.0 to 7.0, and was equivalent to about 36 pct of the inlet sulfur dioxide; the total alkali (fly ash alkali and supplemental lime) was equivalent to about 162 pct of the inlet sulfur dioxide.

At an L/G of 80, inlet level of about 1850 ppm SO_2 (dry), the removal efficiency for the total scrubber system was about 75 pct at the average ash add rate (absorber tower was about 91 pct). Supplemental lime was added to maintain the pH at 6.0 to 7.0, and was chemically equivalent to about 55 pct of the inlet sulfur dioxide; the total alkali (fly ash alkali and supplemental lime) was equivalent to 102 pct of the inlet sulfur dioxide. At the maximum add rate, the removal efficiency for the total scrubber system was about 80 pct (absorber tower about 97 pct), supplemental lime was added to maintain the pH at 6.0 to 7.0, and was chemically equivalent to about 53 pct of the inlet sulfur dioxide; the total alkali was equivalent to about 128 pct of the inlet sulfur dioxide. A test using only lime chemically equivalent to about 134 pct of the inlet sulfur dioxide, the high ash add rate with supplemental lime, resulted in a scrubber system removal of about 80 pct (absorber tower about 97 pct).

The higher sulfur dioxide removals demonstrated in pilot plant tests were obtained by adding lime at rates higher than intended for the 450 MW scrubber unit, and these high rates may not be reproduced in practice on the commercial scale scrubbers. The pilot test results do, however, demonstrate that the scrubber design to be used on the full-scale unit is capable of meeting and exceeding removals required to comply with the 1.2 lb/MM Btu Federal emission standard, and further, that required removals under normal conditions of coal sulfur content can be achieved using fly ash alone without lime.

At the conclusion of the eight-week testing program, the scrubber system was inspected for scale and it was reported to be light, with most deposits at wet-dry interfaces. On the basis of previous pilot plant experience at Montana Power Company, and limited confirmatory testing in this program, it was concluded by CEA-ADL that a suspended solids level of 12 pct provides seed crystals for precipitation of gypsum, and also provides scale control when recirculated to the spray nozzles. However, future reliability tests will be performed to further investigate scaling and corrosion problems. The state of oxidation was high, usually greater than 97 pct. The sludge was reported to have excellent settling properties and cake from the vacuum filter contained 40 to 50 pct moisture.

Operational problems were not numerous and were concerned with an occasional plugging of pipes and erosion of plastic valves and spray nozzles due to the high level of suspended solids.

Burning coal with a sulfur content of about 0.75 pct, CEA-ADL recommends the following full-scale scrubber operating conditions: L/G of 80, 12 pct suspended solids, slurry recycle pH of 5. At the maximum fly ash add rate, no additional alkali would be necessary. Fifteen percent of the incoming flue gas will be by-passed and used for reheat.

Burning coal with the maximum sulfur content (1.3 pct), CEA-ADL recommends the following full-scale operating conditions: A L/G of 80, 12 pct suspended solids, a pH of 6.5 to 6.8, 4 tons per hour of supplemental lime at the maximum fly ash add rate. Fifteen percent of the incoming flue gas will be by-passed and used for reheat.

Capital and operating cost estimates on the 450 MW full-scale scrubber were presented at the Mid-Continent Area Power Pool (MAPP) Environmental Workshop, November 18, 1975[17]. The capital investment estimates for the SBEC project are:

- Material and installation cost	35.5 $/KW
- Additional contracts and site preparation cost	10.0 $/KW
- Escalation, interest during construction, consultant fees, Power Company overhead	20.0 $/KW

The estimated annual operating costs are:

	Unit Cost	¢/MM Btu	$/yr
- Fly ash	0	0	0
- Lime (CaO)	42 $/ton	2.13	$ 555,540
- Waste disposal transportation (30 TPH)	0.80 $/ton	0.99	258,460
- Operating and maintenance staff (20 men)	20,000 $/man	1.23	400,000
- Maintenance (4 pct of capital investment)		1.71	639,000
- Capital fixed charges		6.44	2,092,500
Total Annual Cost		12.5	$3,944,500

If an equivalent lime scrubber were to be constructed, the following operating costs would be affected: an increase of $450,000 per year to purchase 50 tons of lime per day, an increase of about $100,000 per year for offsite disposal of fly ash, a decrease of about $50,000 per year since lower L/G could be employed. The normal operating power requirement for the 450,000 KW fly ash scrubbing system is reported to be 7922 KW. The maintenance and capital fixed charge would be less since it is estimated that the capital investment for the lime system is about 2.5 pct less than the fly ash system. When all economic factors are taken into consideration, the Square Butte Electric Cooperative expects the fly ash alkali scrubbing system to save $444,000 per year in operational costs.

SUMMARY AND CONCLUSIONS OF WESTERN SCRUBBERS

Most future utility scrubbers for Western coals will be designed for the primary purpose of sulfur dioxide removal and not for particulate removal, as was true in the majority of past installations. The major reason is the necessity for having flue gas desulfurization for most low sulfur Western coals under the developing regulatory pattern in the West. There is also a trend toward installing scrubbers in series with electrostatic precipitators on new plants because of the appeal of eliminating the visible fly ash plume with a relatively more reliable device, the ESP. If the scrubber can then be bypassed for short periods for maintenance, a high plant availability can be maintained.

The three most significant factors to be considered in designing scrubbers for Western coals are: 1) the low concentration of sulfur dioxide to be removed, 2) the alkalinity of Western coal fly ashes, and 3) the tendency to operate at a high state of oxidation, producing sulfate and not sulfite. It is significant that all Western FGD installations, existing or planned, use throwaway type processes. This is consistent with the remoteness of Western scrubbers from most large markets for sulfur or gypsum, although sulfuric acid could possibly find sizable markets in the West as an ingredient in fertilizer manufacture. Sulfur dioxide emissions from low-sulfur Western coals can be brought into compliance with the Federal emission standard by treating only a portion of the flue gas and by-passing the remainder for stack gas reheat, as planned by Square Butte Electric Cooperative. Cost is minimized by balancing the savings of treating a smaller volume of gas against the cost of removing a higher percentage of the sulfur dioxide from the fraction treated. Where local standards are more stringent, the option of treating a partial flow ceases to exit.

At low concentrations of sulfur dioxide, it can be argued that gas film diffusion should be the rate controlling step, since the equilibrium partial pressure of sulfur dioxide for alkaline slurry is low and the capacity of the slurry to absorb sulfur dioxide during passage through the scrubber is not taxed if the amount absorbed is small. A sufficient L/G of course plays a part in validating this argument. If gas diffusion does control, design should maximize gas-liquid contact and residence time. Long residence time necessitates a large volume; and good contact requires either multiple sprays or tower packing. If scaling can be resolved, packing is probably the economical choice. If scale has a tendency to form, the large empty volume with multiple sprays will be the better option. The operating experience of the utilities cited in this paper is that, if blowdown is sufficiently restricted, scale will indeed tend to form.

The problem of scaling is inexorably tied in with the question of what constitutes "closed loop" operation. The practical answer is that a system is "closed" if no liquid blowdown is deliberately removed and disposed of. Some inadvertent loss in sludge cannot be eliminated. Beyond this, pond evaporation of a saturated scrubbing liquor does remove sulfate from the system by precipitation, even though liquor may be returned from the pond. Since the sludge and pond losses will vary with design and climate, every system will be "closed" to a different extent.

Control of scaling must be approached differently for low-sulfur Western coals than for Eastern coals. A high state of oxidation and the slight control that can be afforded by manipulating pH in a sulfate system offer little hope that unsaturated operation is possible for Western coals during closed loop operation. That this can be accomplished for high sulfur Eastern coals depends on a low state of oxidation, with consequent precipitation of calcium sulfite and coprecipitation of gypsum from a solution that is not saturated with calcium sulfate[7]. These conditions do not appear to exist for Western coal operations. Control of scale formation, in the authors' opinion, will depend most directly on circulating a sufficiently high level of suspended solids, and operating at a constant pH, whether high or low. This is supported by the 6 to 12 pct level of recirculated suspended solids, and lime or limestone for pH control, currently practiced by the utilities cited in this paper.

Depending on the cost of reagent and the properties of the waste products produced, there may be more or less motivation to improve the utilization of alkalinity in Western fly ashes in scrubbing systems. Laboratory tests at the Grand Forks Energy Research Center have shown that utilization of fly ash alkali improves significantly as the pH is dropped from 5 to 3. Other operating variables would have to be changed along with the pH, including the flow circuit, gas-liquid contact, L/G, and slurry reaction times.

In conclusion, involvement of alkaline ash in a scrubbing system implies a new variable which must be controlled. Since the analyses of Western ashes vary sufficiently, this effectively rules out one tailor-made solution for every scrubber application. As fly ash varies, the characteristics of sludge will vary, and therefore extensive studies on properties of sludge, including leaching of major and trace elements, will be required. Experiences on operating scrubbers indicate that work is needed on materials of construction, and on component design to improve reliability. The solution to stack cleaning problems will require continuing development effort as long as there is a conscious desire for improvement.

APPENDIX

Scrubber Design and Operation
Four Corners Plant
Arizona Public Service Company

LOCATION

1. Farmington, New Mexico.
2. Elevation is 5300 feet.
3. Atmospheric pressure is 12.1 psi.
4. Annual precipitation is 8 inches.
5. Water supply for the plant comes from Morgan Lake, a man-made reservoir filled from the San Juan River.

SCRUBBER APPLICATION

1. Particulate removal, retrofit.
2. Boilers equipped with scrubbers.
    - Two 175 MW Riley pc-fired boilers (Units 1 and 2).
    - One 225 MW Foster Wheeler pc-fired boiler (Unit 3).
3. Service date: Units 1 and 2, December 1971; Unit 3, January 1972.
4. Fuel is New Mexico subbituminous coal from the Navajo mine.
    - 8900 Btu/lb.
    - 12 pct moisture.
    - 0.68 pct sulfur.
    - 22 pct ash.
    - 4 pct CaO in ash.
5. Flue gas entering the scrubbers.
    - Boilers 1 and 2.
        - 814,000 acfm.
        - 340° F.
        - 650 ppm SO_2.
        - 12 gr/scf particulate.
    - Boiler 3.
        - 1,030,000 acfm.
        - Conditions are the same as on units 1 and 2.
7. The particulate removal goal was set by the project at 99.2 pct.

SCRUBBER DESCRIPTION

1. Two venturi scrubbers on each of boilers 1, 2, and 3.
2. Vendor, the Chemico Air Pollution Control Company.
3. Capital cost is $30 million, or $52/kw.
4. Operating costs are not available.
5. Materials of construction.
    - Scrubbers are carbon steel with stainless steel or plastic lining.
    - Outlet ducts were stainless steel lined, later lined with plastic over the stainless steel.
    - Liquid lines and pumps are rubber lined.

- Process vessels are plastic lined.
- Reheaters were 316L stainless steel.
- Wet fans are inconel.

6. No bypass.
7. Turndown is to approximately 50 pct of rated scrubber capacity.
8. Chevron mist eliminators have 6 stages.
9. Wet fan.
10. Reheaters that heated flue gas directly with steam coils failed because of corrosion. The reheat units were removed about one year ago, and no reheat has been used since. Indirect reheat by mixing with heated air is being considered.

SCRUBBER OPERATING DATA

1. L/G is 8.5 gal/1000 acf, or 18 gal/1000 scf.
2. ΔP is 20 to 22 inches H_2O across the venturi, 28 inches overall.
3. "Open loop." Total makeup water for the system is 1700 to 2000 gpm.
4. Gas residence time in the scrubber is not available.
5. Liquid delay time in the venturi recycle loop is about 2 minutes.
6. Liquid temperature leaving the scrubber is 120° F.
7. Solids recirculated has recently been increased from 2 pct to 6 pct.
8. pH of the recycle loop on the scrubber is 3.2 to 3.5. pH at the thickener is 4 to 5 without lime added. A level of pH 7.5 at the thickener is to be maintained with lime addition.
9. Scrubbing liquor analysis is not available.
10. State of oxidation is not available.
11. Degree of supersaturation is not available.

OPERATING REQUIREMENTS

1. Lime is added at the rate of 6 tons/day for pH control.
2. No dispersing agent is being used.
3. System makeup water requirements are about 3400 acre ft/yr.
4. Power requirements.
   - Electrical requirements are 3 to 4 pct of generating capacity.
5. Manpower.
   - 8 operators.
   - Maintenance personnel not available.

OPERATING RESULTS

1. Particulate removal meets the goal of 99.2 pct.
2. SO_2 removal is 30 to 35 pct without lime.
   - No typical SO_2 removal has been determined with lime.
3. Availability overall is estimated at 80 pct.
4. Scaling has occurred on most wetted surfaces.
5. Methods used for scale control.
   - The level of recirculated solids was recently increased from 2 to 6 pct.
   - A lime system has recently been installed to maintain pH at 7.5 in the thickener.
   - The amount of blowdown used helps to control scaling.

6. Problems.
   - The principal problem is that scaling is not under control. The effects of high pH in the thickener and 6 pct solids in the recirculation scrubber liquor have not yet been assessed because of their recent implementation, and a long-range testing program is planned.
   - Corrosion with resulting leakage has occurred where coating on mild steel have failed.
   - Solids buildup has occurred in blowdown lines.
   - Deterioration of stack linings.
7. Disposal of sludge.
   - Sludge settles well without a flocculating agent. The sludge is concentrated to 30 or 35 pct solids in the thickener underflow and is pumped after some dilution with blowdown to decanting ponds. Ash at present is left to accumulate in the ponds, but it may be dredged and returned to the mine.

Scrubber Design and Operation
Aurora Station
Minnesota Power and Light

LOCATION

1. Aurora, Minnesota.
2. Elevation 1500 feet.
3. Atmospheric pressure 14.1 psi.
4. Annual rainfall approximately 25 inches.

SCRUBBER APPLICATION

1. Particulate removal, retrofit.
2. Two 58-MW pc-fired boilers.
3. Scrubber startup, June 1971.
4. Fuel is Montana subbituminous coal from the Big Sky Mine.
   - 8800 Btu/lb.
   - 26.7 pct moisture.
   - 1.37 pct sulfur (mine average).
   - 9 pct ash.
   - 9 to 13 pct CaO in ash.
5. Flue gas entering the scrubber on each boiler.
   - 291,160 acfm.
   - 340° F.
   - 800 ppm SO_2.
   - 2.06 gr/scf.
6. Removal goals.
   - Minnesota particulate standard is 0.6 lb/MM Btu.
   - Scrubber guarantee is 0.03 gr/scf or 0.078 lb/MM Btu.

SCRUBBER DESCRIPTION

1. One Elbair spray-impingement scrubber for each of the two boilers.
2. One-stage of high pressure spray is directed concurrent with gas flow against vertical rods to be atomized into fine droplets.

3. Vendor, Krebs Engineers.
4. Size of each scrubber is nominally 60 MW or 300,000 acfm.
5. Capital cost is not available.
6. Operating cost is not available.
7. Materials of construction.
   - Scrubber is 316 ELC stainless steel.
   - Outlet ducts are flake-polyester coated carbon steel.
   - Piping is fiberglass, or rubber lined.
   - Pumps are rubber lined.
8. No bypass.
9. Turndown is 0 to 110 pct.
10. Demistor consists of one bank of vertical chevrons.
11. No reheat, wet fan.

SCRUBBER OPERATING DATA

1. L/G is 8.3 gal/1000 acf, or 13.3 gal/1000 scf.
2. ΔP is 2.5 inches of H_2O across scrubber, 4 inches total.
3. Not closed loop, although scrubbing liquor is recycled from the ash pond back to the scrubber. No clarifier is used. Makeup water is approximately 1200 gpm for each unit.
4. Gas residence time in the scrubber is approximately 3 seconds.
5. Liquid delay time in the recycle circuit is not available.
6. Solids circulated.
   - 0.02 pct entering scrubber.
   - 0.75 pct leaving scrubber.
7. pH estimated at 4.5 in, 4.4 out.
8. Scrubbing liquor analysis is not available.
9. State of oxidation of dissolved sulfur is high, estimted over 90 pct sulfate.
10. Saturation of scrubbing liquor does not occur at the level of blowdown used.

OPERATING REQUIREMENTS

1. No reagent is used.
2. Scrubber water requirements are about 3500 acre ft/yr.
3. Power requirement.
   - Electrical power is about 0.5 MW per unit, or 0.8 pct.
   - No steam is used for reheat.
4. Manpower requirements are not available.

OPERATING RESULTS

1. Particulate removal is about 98 pct; exit dust loading is 0.04 to 0.046 gr/scf.
2. SO_2 removal, occurring incidental to particulate removal by reaction with the alkaline fly ash and by sulfate removal in blowdown, is typically 20 pct.
3. Availability.
   - The Elbair scrubber, consisting of a stainless steel box containing high pressure spray nozzles that are removeable section by section for maintenance, can remain on line without a bypass, even though the spray system is not operating. However, effective operation requires a high level of maintenance effort. A percentage availability in terms of effective operation was not obtainable.

4. Scaling and plugging.
   - Scaling and plugging has not been too severe, owing to the amount of blowdown used.
5. Disposal of ash and spent scrubbing liquor.
   - Blowdown, estimated to contain 1 pct solid fly ash, is sent an ash pond. Overflow from the pond is neutralized with lime before disposal.
6. Problems.
   - This unit would suffer from the same scaling and plugging problems as the Clay Boswell scrubber of similar design, described in the preceding section, if the recycle loop were closed to a similar extent. As operated, it is less troublesome.
   - The mist carryover problem is less severe than at Clay Boswell owing to operation at partial load with a resultant lower stack velocity.

Scrubber Design and Operation
Clay Boswell Plant
Minnesota Power and Light

LOCATION

1. Cohasset, Minnesota.
2. Elevation approximately 700 feet.
3. Atmospheric pressure 14.3 psi.
4. Annual rainfall approximately 25 inches.

SCRUBBER APPLICATION

1. Particulate removal for a new plant.
2. 350 MW Combustion Engineering pc-fired boiler.
3. Scrubber startup May 1973.
4. Fuel is Montana subbituminous coal from the Big Sky Mine.
   - 8800 Btu/lb.
   - 26.5 pct moisture.
   - 1.37 pct sulfur (mine average)
   - 9 pct ash.
   - 9 to 13 pct CaO in the ash.
5. Flue gas entering the scrubber.
   - 1,300,000 acfm.
   - 254° F.
   - 800 ppm SO_2.
   - 3 gr/scf.
6. Removal goals.
   - Minnesota particulate standard is 0.6 lb/MM Btu.
   - Scrubber guarantee is 0.03 gr/scf or 0.078 lb/MM Btu.

SCRUBBER DESCRIPTION

1. A single Elbair spray-impingement scrubber.
2. One stage of high pressure spray is directed concurrent with gas flow against punch plate baffles to be atomized into fine droplets.
3. Vendor, Krebs Engineers.
4. Size of scrubber is nominally 350 MW or 1.3×10^6 acfm.
5. Capital cost is not available.
6. Operating cost is not available.
7. Materials of construction.
   - Scrubber is 316 LC stainless steel.
   - Outlet ducts are flake-polyester coated carbon steel.
   - Piping is fiberglass or rubber lined.
   - Pumps are rubber lined.
8. No bypass.
9. Turndown is 0 to 110 pct.
10. Demistor consists of one bank of vertical chevrons.
11. No reheat, wet fan.

SCRUBBER OPERATING DATA

1. L/G -- 8.3 gal/1000 acf, or 13.3 gal/1000 scf.
2. ΔP is 2.4 inches H_2O across scrubber, 4 inches total.
3. Not "closed loop" although scrubbing liquor is recycled from the clarifier back to the scrubber. Makeup water is approximately 1400 gpm. About 420 gpm of the makeup water compensates for clarifier underflow to the fly ash pond. The remainder is clarifier overflow which is pumped to the bottom ash pond. The clarifier overflow blowdown maintains the sulfur level at about 3000 ppm.
4. Gas residence time in the scrubber is 3 seconds.
5. Liquid delay time in one clarifier is 2 hours, or 4 hours if two clarifiers are on line.
6. Solids circulated.
   - 0.02 pct entering scrubber.
   - 0.75 pct leaving the scrubber.
7. pH 4.5 in, 4.0 out.
8. Scrubbing liquor analysis.
   - Ca -- 600 ppm.
   - Mg -- 200 ppm.
   - Na -- 15 ppm.
   - SO_4 -- 2300 ppm.
9. State of oxidation of dissolved sulfur is high, estimated over 90 pct sulfate.
10. Degree of supersaturation is not measured, but it is evident that it is variable depending on the amount of soluble alkali in the ash, the amount of SO_2 absorbed, and the amount of blowdown removed from the system.

OPERATING REQUIREMENTS

1. No reagent is regularly used.
2. Water requirements are about 2300 acre ft/yr.

3. Power requirement.
   - Electrical power is about 3 MW, or 0.86 pct of net generating capacity.
   - No steam is used for reheat.
4. Manpower requirements are not available, but maintenance is known to be very high because of a continuous schedule of cleaning on the mist eliminators, which require sand blasting to clean. As of July 1974, 85 to 100 manhours were spent each week removing fly ash and calcium sulfate scale deposits. In the past, the spray nozzles also required continuous maintenance, however, with the increased blowdown now employed, the nozzles require only periodic maintenance.

OPERATING RESULTS

1. Particulate removal is about 97 pct; exit dust loading is 0.08 pct gr/scf.
2. SO_2 removal, occurring incidental to particulate removal by reaction wit the alkaline fly ash and by sulfate removal in blowdown, is typic[illegible] 15 to 20 pct.
3. Availability.
   - The Elbair scrubber, consisting of a stainless steel box containing high pressure spray nozzles that are removable for maintenance section by section, can remain on line without a bypass, even though the spray system is not operating. However, effective operation requires a very high level of maintenance effort. A percentage availability in terms of effective operation was not obtainable.
4. Scaling and plugging.
   - Scaling and plugging occur in nozzles, nozzle trees, strainers, on punch plate baffles, in the wet-dry zone, and on the fan and mist eliminator; and deposits fall into the drains at the bottom of the scrubber.
   - Scaling is aggravated by any increase in CaO content in the ash of the coal being burned, which causes increased SO_2 removal and an increase in the level of Ca^{++} and $SO_4^{=}$ ions in solution.
5. Measures used for scale control.
   - Substantial amounts of blowdown are removed from the system to remain below saturation with $CaSO_4$.
   - A very high level of cleaning and maintenance is carried out continuously. Nozzles and nozzle trees are removed and cleaned on a rotating schedule, with all nozzles being cleaned once per week.
6. Disposal of ash and spent scrubbing liquor.
   - Blowdown from two clarifiers, containing typically 5 or 6 pct fly ash, is sent to an 80-acre ash pond. There is no net evaporation, but rather an accumulation of about 10 inches of water per annum.
7. Problems.
   - No ultimate solution to the problem of disposing of sulfate-laden blowdown water has been found. The two possible solutions would involve either unrestricted discharge of diluted blowdown to streams or closing the loop in the scrubber circuit to eliminate blowdown. A break-through in methods of scale control would be required to eliminate blowdown.

- A stack mist problem results primarily from washing of the wet fan to remove scale buildup. Attempts to use steam soot blowers in place of washing and to apply non-stick coating have been unsuccessful. Some improvement has been achieved by reducing the amount of fan washing. Recent tests show favorable results by using low-velocity stacks.

8. Because of the difficulties experienced with flue gas mist carryover, scale deposits on the mist eliminators and ID fan, disposal of sulfate-laden water and the high maintenance rate, the construction of similar scrubbers for the two 70 MW boilers was not completed. In lieu of the particulate scrubbers, ESP's are presently planned. Flue gas from all three units would be discharged from one low velocity stack and the two 70 MW boilers should provide reheat for flue gas exiting the 350 MW scrubber.

Scrubber Design and Operation
Dave Johnston Plant
Pacific Power and Light

LOCATION

1. Glenrock, Wyoming
2. Elevation approximately 5000 feet.
3. Atmospheric pressure 12.3 psi.
4. Annual precipitation is 14 inches.
5. Plant water supply is the North Platte River.

SCRUBBER APPLICATION

1. Particulate removal, retrofit.
2. One 330 MW Combustion Engineering pc-fired boiler (Unit no. 4).
3. Three parallel scrubbers.
4. Scrubber startup was April 1972.
5. Fuel is Wyoming subbituminous coal from a captive mine. Typical analysis is:
    - 7430 Btu/lb.
    - 26 pct moisture.
    - 0.5 pct sulfur.
    - 12 pct ash.
    - 20 pct CaO in ash.
6. Flue gas entering scrubber.
    - 1,500,000 acfm.
    - 270° F.
    - 500 ppm SO_2.
    - 12 gr/scf (design).
    - 4 gr/scf (actual).
7. Removal goal.
    - 99.7 pct removal, or 0.04 gr/scf exit dust loading.

SCRUBBER DESCRIPTION

1. Three identical scrubbers in parallel.
2. Venturi scrubbers.
3. Vendor, the Chemical Construction Company (Chemico).
4. Initial capital investment was $8 million, $24/kw. Costs incurred since startup have increased this amount significantly.
5. Operating cost is not available.
6. Materials of construction.
   - Scrubbers, vessels, outlet duct, and stack are polyester lined steel.
   - Piping and fan housing are rubber lined.
   - Fan wheels are Inconel.
7. No bypass.
8. Turndown is to approximately 30 pct of rated scrubber capacity.
9. Chevron mist eliminator.
10. Wet fans, no reheat.

SCRUBBER OPERATING DATA

1. L/G is 13.3 gal/1000 acf, or 22 gal/1000 scf.
2. ΔP is 10 inches of H_2O across the venturi, 15 inches total.
3. Intermittently "open loop." Normal operation is attempted at a makeup rate of 500 gpm, which compensates for evaporation and loss in sludge. The unit has been operated at times with 3000 gpm fresh water makeup to flush out scale.
4. Gas residence time in the venturi section of the scrubber is estimated at about 1 second.
5. Liquid exit temperature is 126° F.
6. Liquid delay time in the venturi recycle loop is 2 to 3 minutes.
7. Solids recirculated are 2 pct of scrubbing liquor.
8. pH leaving the scrubber is about 5.
   Future tests will be conducted maintaining the pH at 5.5 to 6.0 using hydrated lime.
9. Scrubbing liquor analysis is not available.
10. Degree of supersaturation is from 1.0 to 1.3.

OPERATING REQUIREMENTS

1. Lime is added for pH control.
2. Lignosulfonate is added to minimize scale at wet-dry zones.
3. Water requirements are approximately 800 acre-ft/yr in "closed loop" mode. Actual requirements are greater because of occasional flushing.
4. Power requirements.
   - Electrical power requirement is 7 to 8 MW, or 2.3 pct of generating capacity.
   - No steam is used for reheat.
5. Manpower requirements are not available.

OPERATING RESULTS

1. Particulate removal meets the outlet grain loading goal of 0.4 gr/scf.
2. SO_2 removal (preliminary values).
   - 35 to 40 pct without lime.
   - Lime addition results in modest increase in SO_2 removal. Exact value has not been determined.
3. Availability not available, but it is not considered adequate for a utility power source.
4. Scaling and plugging.
   - Solids buildup has occurred at the wet-dry interface.
   - Hard gypsum scale has formed in the scrubber vessels and piping.
   - Solids plug bleed and recycle lines.
5. Methods for controlling scaling and plugging.
   - Lime for pH control has resulted in reduction but not elimination of scaling.
   - Lignosulfonate addition has resulted in a less adherent or friable wet-dry buildup.
   - Effects of hexamataphosphate on scaling and wet-dry buildup were determined to be ineffective and its use discontinued.
   - Continuous fan wash has essentially eliminated buildup on fans. The wash water is composed of cooling water blowdown and service water.
   - Fresh water washing has been required to flush ash and scale deposits from the scrubber vessel.
6. Additional problems.
   - Recycle pump erosion.
   - "Silting" during shutdown.
7. Disposal of sludge.
   - Bleed from the scrubber circuit is sent directly to two ash ponds, from which overflow flows to a clear pond for recycle to the scrubber circuit. Each ash pond is dredged once per year, and the solids hauled out for landfill. Excess water resulting from periods of flushing is discharged to the North Platte River under a variance from the State of Wyoming.

Scrubber Design and Operation
Valmont, Cherokee, and Arapahoe Stations
Public Service Company of Colorado

NOTE: Public Service Company of Colorado has five TCA scrubbers containing thirteen modules installed on five boilers at three stations. Because of the close similarity between these installations, they will be discussed collectively rather than individually.

LOCATIONS

	Valmont Boulder, Colo.	Cherokee North Denver, Colo.	Arapahoe Southwest Denver, Colo.
Elevation, ft	5300	Est. 5200	Est. 5600
Atm. P, psi	12.1	12.1	12.1
Annual rainfall, in.	14	14	14
Plant water supply	Hillcrest Lake	South Platte River	South Platte River

SCRUBBER APPLICATION

1. All units are retrofits for particulate removal; Valmont #5 has one module modified to study SO_2 removal.
2. Boilers equipped with scrubbers (all pc-fired).
3. Valmont #5, 196 MW, 2 scrubber modules, November 1971.
   Cherokee #1, 115 MW, 1 scrubber vessel, 2 modules, June 1973.
   Cherokee #3, 170 MW, 1 scrubber vessel, 3 modules, November 1972.
   Cherokee #4, 375 MW, 1 scrubber vessel, 4 modules, July 1974.
   Arapahoe #4, 112 MW, 1 scrubber vessel, 2 modules, September 1973.
4. Arrangements of particulate cleaning equipment.
   - At Valmont, flue gas from a mechanical collector is split into two parallel streams, with 50 pct sent to the scrubber and 50 pct to an electrostatic precipitator (ESP).
   - All other units have a mechanical collector, an ESP, and scrubber(s) in series, with all flue gas entering the scrubber(s).
5. Coal burned at Valmont and Arapahoe is Wyoming subbituminous.
   - 8300 Btu/lb.
   - 29 pct moisture.
   - 0.6 pct sulfur.
   - 5.2 pct ash.
   - 20 pct CaO in ash.
6. Coal burned at Cherokee is Colorado bituminous coal.
   - 11,000 Btu/lb.
   - 9.8 pct moisture.
   - 0.7 pct sulfur.
   - 9.4 pct ash.
   - 5 pct CaO in ash.
7. Flue gas entering the stack gas cleaning train.

	Valmont #5	Cherokee #1	Cherokee #3	Cherokee #4	Arapahoe #4
Flow, acfm	463,000	520,000	610,000	1,520,000	520,000
T, °F	260	295	280	275	300
SO_2, ppm (est.)	500	500	500	500	500
Dust, gr/scfd	0.8	0.8	0.4	0.7	0.8

8. Removal goals for particulate.
   - The applicable Colorado State Standard is 0.1 lb/MM Btu, or about 0.05 gr/scf.
   - The company's desire for clean stacks requires a goal of 0.02 gr/scf.

SCRUBBER DESCRIPTION

1. All units are Turbulent Contact Absorbers, consisting of three stages of mobile packing, or "ping pong balls," with spray directed downward through the balls and gas passing countercurrent upward.
2. Vendor: Air Correction Division, Universal Oil Products Company.
3. Capital cost:

Valmont #5	Cherokee #1	Cherokee #3	Cherokee #4	Arapahoe #4
#3,600,000	$3,810,000	$4,400,000	$12,700,000	$4,560,000
$32/kw	$33/kw	$29/kw	$33/kw	$41/kw

4. Operating costs for Cherokee #3 is 0.50 mills/kwh. Other operating costs are not available.
5. Materials of construction (typical).
   - Scrubbers are rubber-lined carbon steel with stainless steel grids.
   - Exit ducts are mild steel with stainless steel and/or fabric expansion joints.
   - Slurry piping is rubber lined.
   - Pumps are rubber lined.
6. Bypasses on all units.
7. Typical turndown is 47 to 105 pct.
8. Mist eliminators have 2 stages, 7 passes.
9. Demisters for Cherokee #1 and #3 are fiberglass reinforced plastic. Demisters for Cherokee #4, Arapahoe #4 and Valmont #5 are stainless steel.
10. All units except Cherokee #4 reheat the flue gas directly with steam coils; Cherokee #4 uses externally-heated air.
11. All units have dry fans that are forced-draft with respect to the scrubber.

SCRUBBER OPERATING DATA FOR CLOSED LOOP MODULE AT VALMONT #5

1. L/G of about 50.
2. ΔP is approximately 10 to 15 inches of H_2O.
   - Three stages of mobile packing (8-12 inches per stage).
   - Stainless steel mist eliminators.
   - Direct steam coil reheater, 1 1/2 to 4 inches.
   - Transition ductwork, 1/2 inch.
   - Total, 16 1/2 inches maximum.
3. Makeup water is 70 gpm. The water analysis is as follows:
   - SO_4 -- 45 ppm.
   - Ca -- 35 ppm.
   - Cl -- 10 ppm.
4. Gas residence time is not known.
5. pH entering is maintained at 5.5 to 6.5 by limestone addition.
6. State of oxidation of dissolved sulfur is nearly 100 pct sulfate.
7. Suspended solids maintained at 7 pct.
8. Liquid temperature is 110° F.
9. Degree of supersaturation is not known.

SCRUBBER OPERATING DATA FOR OPEN LOOP SYSTEMS

1. L/G is typically 54-59.
2. ΔP is approximately 10 inches to 15 inches of H_2O depending on design and operating conditions.
   - 3 stages of mobile packing -- 8-12 inches.
   - Mist eliminators.
     - Fiberglass reinforced plastic, 1 1/2 to 3 inches.
     - Stainless steel -- 1/2 inch.
   - Reheat.
     - Direct steam coil reheater, 1 1/2 to 4 inches.
     - Hot air -- negligible.
   - Transition ductwork -- 1/2 inch
   - Total 16 1/2 inches maximum
3. "Open loop." Amounts of makeup water from cooling tower blowdown are as follows:

Cherokee #1	Cherokee #3	Cherokee #4	Arapahoe #4	Valmont #5 (unmodified module)
203 gpm	380 gpm	744 gpm	203 gpm	210 gpm

4. Gas residence time in the scrubber is 3.8 to 5 seconds.
5. Liquid temperature leaving the scrubber is 110° F.
6. Liquid holdup time in the recycle circuit is very short, estimated at ten seconds.
7. pH is 7 to 9 entering; 2.8 to 3 leaving the scrubber.
8. A scrubber liquor analysis for Valmont (not known to be representative).
   - Ca -- 590 ppm.
   - Mg -- 350
   - Na -- 1
   - SO_4 -- 10,000
9. State of oxidation of dissolved sulfur is not available.
10. Degree of supersaturation is not available.

OPERATING REQUIREMENTS

1. No lime or other reagent or additive is normally used.
2. Water requirements in acre-ft/yr (approximate).

Valmont #5	Cherokee #1	Cherokee #3	Cherokee #4	Arapahoe #4
340	327	612	1200	327

3. Power requirements.

Electric: Power:

Valmont #5	Cherokee #1	Cherokee #3	Cherokee #4	Arapahoe #4
5.3 MW	5.14 MW	6.4 MW	14.4 MW	5.2 MW
5.4 %	4.5 %	3.8 %	3.8 %	4.6 %

Steam for reheat:

	Valmont #5	Cherokee #1	Cherokee #3	Cherokee #4	Arapahoe #4
lb/hr:	50,000	50,000	41,200	135,000	60,000
°F:	700	420	715	634	360
psi:	490	300	300	1,975	150

4. Manpower for scrubber operation is not identified by the company as a separate category from plant oeprators. However, it is reported that a high degree of maintenance is required.

OPERATING RESULTS

1. Particulate removal achieves an outlet grain loading of 0.02 gr/scf.
2. SO_2 removal.
   - 45 to 50 pct at Valmont and Arapahoe, burning Wyoming coal.
   - 15 to 20 pct at Cherokee, burning Colorado coal.
   - 85 pct at Valmont using modified module (not used continuously).
3. Availability.
   - Valmont #5 -- 55 pct for 46 months of operation.
   - Cherokee #1 -- 53 pct for 23 months of operation.
   - Cherokee #3 -- 66 pct for 35 months of operation.
   - Cherokee #4 -- 82 pct for 10 months of operation.
   - Arapahoe #4 -- 84 pct for 25 months of operation.
4. Problems.
   - Scaling and plugging has occurred at:
     - The wet/dry zone.
     - The first stage grid.
     - The reheater.
     - The mist eliminators.
   - Corrosion has caused major failures of reheaters at the Cherokee Station.
   - Wear on the balls used for packing requires replacement after 6000 hours or less.
   - The cost of replacing the balls are 0.06 $/ball. The number of balls required is as follows:
     - Valmont #5 -- 870,000 balls.
     - Cherokee #1 -- 980,000 balls.
     - Cherokee #3 -- 1,180,000 balls.
     - Cherokee #4 -- 3,070,000 balls.
     - Arapahoe #4 -- 980,000 balls.
5. Measures for control of scaling.
   - Additives tried have not worked, including phosphated esters. The studies were limited in that they were of relatively short duration.
   - Blowdown must be maintained at an adequately high level, but otherwise no specific methods are being used.
   - One of two modules at the Valmont Station has been modified to use limestone for pH control. The preliminary results are favorable: however, additional tests are planned for the future.

6. Sludge disposal.
   - At the Arapahoe Station slurry blowdown is pumped to an ash settling pond after adjusting the pH to 6.0 - 9.0 with lime. The settled fly ash is dredged periodically and used for landfill. Clear effluent from the ponds is discharged under permit from the State of Colorado, to the South Platte River at the Cherokee and Arapahoe Stations, and to the cooling pond at the Valmont Station. At the Cherokee Station only slurry blowdown is neutralized and clarified and then mixed with ash pond overflow for discharge to the river, owing to the heavy loading on the ponds caused by discharge from the three scrubber-equipped boilers at this station.

ACKNOWLEDGMENTS

Information contained in this paper was obtained from published sources and, from inquiries directed to utilities operating wet scrubbers on boilers burning Western coals and to scrubber vendors. Persons who have contributed information include Dr. Fred Murad of Combustion Equipment Associates, Mr. Ken Vig of Minnkota Power Cooperative, Mr. Dennis Van Tassel and Mr. Eldon Kilpatrick of Minnesota Power and Light, Mr. Phil Richmond of Square Butte Electric Cooperative, Mr. George Greene and Mr. Steven Goering of Public Service of Colorado, Mr. Thomas Ashton of Pacific Power and Light Company, Dr. Carlton Grimm of Montana Power Company, Mr. John Noer of Northern States Power, Mr. Walter Ekstrom and Mr. Aubrey Parsons of Arizona Public Service. The contributions of the above and others who have freely exchanged information in numerous past contacts concerning flue gas desulfurization are gratefully acknowledged.

REFERENCES

1. U.S. Bureau of Mines, Division of Fossil Fuels. Coal--Bituminous and Lignite in 1973. Mineral Industry Surveys, January 4, 1975, p 5.

2. Nielsen, G.F. Coal Mine Development Survey. Coal Age. V. 80, February 1975, pp 130-139.

3. Energy Research and Development Administration. Open file report. Survey of Coal and Ash Composition and Characteristics of Western Coals and Lignite. Grand Forks, ND, 1975.

4. Gronhovd, G.H., et al. Some Studies on Stack Emissions from Lignite Fired Power Plants. BuMines IC 8650, 1974, pp 103, 133.

5. Facts Sheet. Four Corners Powerplant, Farmington, NM, April 1973.

6. Quig, R.H. Chemico Experience for SO_2 Emission Control on Coal-Fired Boilers. Presented at the Coal and the Environment Technical Conference, Louisville, KY. October 23, 1974.

7. Borgwardt, R.H. EPA/RTP Pilot Studies Related to Unsaturated Operation on Lime and Limestone Scrubbers.

8. Tufte, P.H., et al. Pilot Plant Scrubber Tests to Remove SO_2 Using Soluble Alkali in Western Coal Ash. BuMines IC 8650, 1974, pp 103-133.

9. Sondreal, E.A., et al. Wet Scrubbing of SO_2 with Alkali in Western Coal Ash. Paper No. 74-272, 67th Annual Meeting of the Air Pollution Control Association, June 9-13, 1974, 31 pp.

10. Sondreal, E.A., and P.H. Tufte. Scrubber Developments in the West. Presented at the Lignite Symposium, Grand Forks, North Dakota, May 14-15, 1975.

11. LaMantia, C., et al. EPA-ADL Dual Alkali Program Interim Results. Presented at EPA Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.

12. Kilpatrick, E.R., and H.E. Bacon. Experience with a Flue Gas Scrubber on Boilers Burning Subbituminous Coal. American Society of Mechanical Engineers Winter Annual Meeting, New York, NY, November 1974. Paper No. 74-WA/APC-3.

13. LaMantia, C.R., and I.A. Raben. Some Alternatives for SO_2 Control. Presented at Coal and the Environment, Technical Conference sponsored by the National Coal Association, October 22-24, 1974.

14. Noer, J.A., et al. Results of a Prototype Scrubber Program for the Sherburne County Generating Plant. Presented at the IEEE-ASME Joint Power Generation Conference, Miami Beach, Florida, September 15-19, 1974.

15. Ashton, T.M. Operating Experience Report, Flue Gas Scrubbing System, Dave Johnston Steam-Electric Plant Unit 4, Pacific Power and Light Company, presented at the American Society of Mechanical Engineers National Symposium, Philadelphia, PA., April 1973.

16. Green, G.P. Operating Experience with Particulate Control Devices. Presented at the American Society of Mechanical Engineers National Symposium, Philadelphia, PA, April 1973.

17. Murad, F.Y., et al. Boiler Flue Gas Desulfurization by Fly Ash Alkali. Presented at Mid-Continent Area Power Pool (MAPP) Environmental Workshop, Minneapolis, MN, November 18, 1975.

RESULTS OF THE 170 MW TEST MODULES PROGRAM
MOHAVE GENERATING STATION
SOUTHERN CALIFORNIA EDISON COMPANY

Alexander Weir, Jr., Lawrence T. Papay, Dale G. Jones,
John M. Johnson, and William C. Martin

Southern California Edison Company
P. O. Box 800
2244 Walnut Grove Avenue
Rosemead, California 91770

ABSTRACT

This paper summarizes the performance of three different types of scrubbers tested with lime and limestone reagents at the Mohave Generating Station in South Point, Nevada. Each scrubber was designed to treat 450,000 SCFM of flue gas (170 megawatt equivalent) and was larger than any other single scrubbing module which has been operated in the world today. The Horizontal Module (a horizontal cross flow spray scrubber) was operated from January 16, 1974 to February 9, 1975. The Vertical Module was tested in two modes; first as a Turbulent Contact Absorber (TCA) from November 2, 1974 to April 30, 1975 when the thermoplastic "ping pong balls" were removed and second as a Polygrid Packed Absorber (PPA), with a plastic "eggcrate" packing with testing continuing to July 2, 1975.

The effects of recirculating slurry flow rate, flue gas flow rate (turndown) and number of scrubbing stages on SO_2 and particulate removal are presented. Performance of two types of flue gas reheaters and two types of mist eliminators are discussed. Power requirements, reagent utilization factors, scrubbing system, availability records and closed loop water system operations are described for three types of scrubbing systems tested at Mohave.

FOREWORD

The Test Modules Program was a joint venture of the Navajo and Mohave Power Project participants who are listed below:

- Salt River Project Agriculture Improvement and Power District
- Arizona Public Service Company
- Department of Water and Power of the City of Los Angeles
- Nevada Power Company
- Tucson Gas and Electric Company
- Bureau of Reclamation of the U.S. Department of the Interior
- Southern California Edison Company

Funding for this program was provided by the participants in accordance with their respective megawatt entitlements in the Navajo and Mohave Power Projects. Southern California Edison Company was the project manager of the Test Modules Program.

The conclusions presented in this paper represent the personal opinions of the authors and are not intended to represent the opinions or position of any of the project participants.

INTRODUCTION

At the second EPA Flue Gas Desulfurization Symposium in New Orleans in 1972, Dr. Jerry Shapiro[1] of the Bechtel Power Corporation presented the plans of the Navajo and Mohave participants to test, on a pilot plant scale, four different types of scrubbers with three different reagents (lime, limestone, and soda ash). The results of these tests, as well as tests of four additional scrubbers with one additional reagent (ammonia), were presented at the third Flue Gas Desulfurization Symposium in New Orleans in May 1973.[2] At that time it was indicated that contracts had been let in December of 1972 to construct two 170 MW Test Modules at the Mohave Generating Station. Both of these scrubbers were larger than any other single scrubbing module which had been previously operated in the United States, or in the world. With a rated capacity of 450,000 SCFM, that statement is still true today. Testing of the Horizontal Module was initiated January 16, 1974 and the results of the first nine months of operation were presented at the fourth Flue Gas Desulfurization Symposium in Atlanta on November 4, 1974. Testing of this module was completed February 9, 1975 after 5927 hours of operation and it has since been dismantled and reassembled at the Four Corners Generating Station.

Start-up of the Vertical Module was initiated on schedule January 1, 1974, but on January 24, 1974 a disasterous fire burned most of the chlorobutyl rubber lining and start-up was not resumed until October 1, 1974.[3] Testing of the Vertical Module was initiated November 1, 1974, and was completed July 1, 1975 after 3131 hours of test. Two different configurations of the Vertical Module were tested. The first configuration was the TCA (Turbulent Contact Absorber) configuration in which four stages and later three stages of thermoplastic rubber balls were used. Later, the balls were removed and replaced with a plastic "eggcrate" type packing referred to in this paper as the PPA (Polygrid Packed Absorber) configuration. The Vertical Module is presently shut down in a cold standby condition at the Mohave Generating Station.

It is the purpose of this paper to present the results of testing of these three types of scrubbers at the 170 MW scale, as well as results of testing of two types of reheaters, and two types of demisters. Not included in this paper, for lack of time, were the results of testing various scrubber lining materials. Six types of slurry pumps, as well as rotary filters, centrifuges and two commercially available methods of sludge fixation (as well as sludge ponding) were also evaluated in this program but these results also will not be presented in this paper. Since another session of this symposium is directed to slurry and sludge handling, EPA has requested that we not discuss this aspect of a total scrubbing system in this paper.

TEST MODULE DESCRIPTIONS

170 MW Vertical TCA and PPA

A ground-level photograph of the Vertical Module is shown in Figure 1, where the flue gas enters the scrubbing chamber at the lower left and exits from the top of the scrubber. Although this photograph does not show the front end limestone preparation area, back-end thickener nor sludge dewatering equipment, it does indicate the height to the top of the outlet ductwork, which is about 145 feet above grade. The ductwork shown at the left side of Figure 1 supplies flue gas to the scrubber from the down stream side of the Unit 1 electrostatic precipitator. A 5,500 HP booster fan forces the flue gas through the inlet ducting, scrubber packing and mist eliminator section into the outlet ducting where it is heated by direct-contact steam coils and is finally returned to the ductwork leading to the 500 foot stack.

The majority (2,342 hours) of the Vertical Module test program was conducted with the Turbulent Contacting Absorber (TCA) configuration shown in Figure 2. The scrubbing chamber dimensions are 18 feet wide by 40 feet long and 38 feet high. The TCA configuration consisted of four stages of thermoplastic rubber balls supported on stainless steel grids at four foot intervals. The balls were contained in compartments, with 15 compartments at each of the four levels. Although various levels of ball depths were tested, the compartments were initially filled to the one-foot level with approximately 1,600,000 balls. The TCA configuration was tested with both three and four stages of balls and at ball depths of 6, 10 and 12 inches in the four stage configuration.

During that last phase of the program, the TCA configuration was replaced with a Polygrid Packed Absorber (PPA), as shown in Figure 3. The Vertical Module was operated for 789 hours in the PPA configuration. The Polygrid packing consisted of plastic grids in an "eggcrate" configuration where each grid layer was 1 1/4" thick with 2" square openings. The grid layers were stacked to a depth of approximately 17 inches in each stage.

The Vertical Module recirculating slurry flow rate was normally 16,000 gpm in the TCA configuration and 27,000 gpm in the PPA configuration. A small quantity of the recirculating slurry was discharged to a dewatering complex to extract solids from the scrubber slurry. Water reclaimed in the dewatering process was returned to the limestone slurry mix tank. The dewatering complex consisted of a thickener tank, with further dewatering provided by either a centrifuge or vacuum drum filters.

170 Horizontal Module

The photograph shown in Figure 4 shows the size and general arrangement of the basic scrubber shell, ductwork and major components of the Horizontal Module. The horizontal end-to-end length including the booster fan, demister section and ductwork is 150 feet long. The width varies from 28 to 32 feet and the top of the scrubber shell is 33 feet above grade. The ductwork shown on the right side of Figure

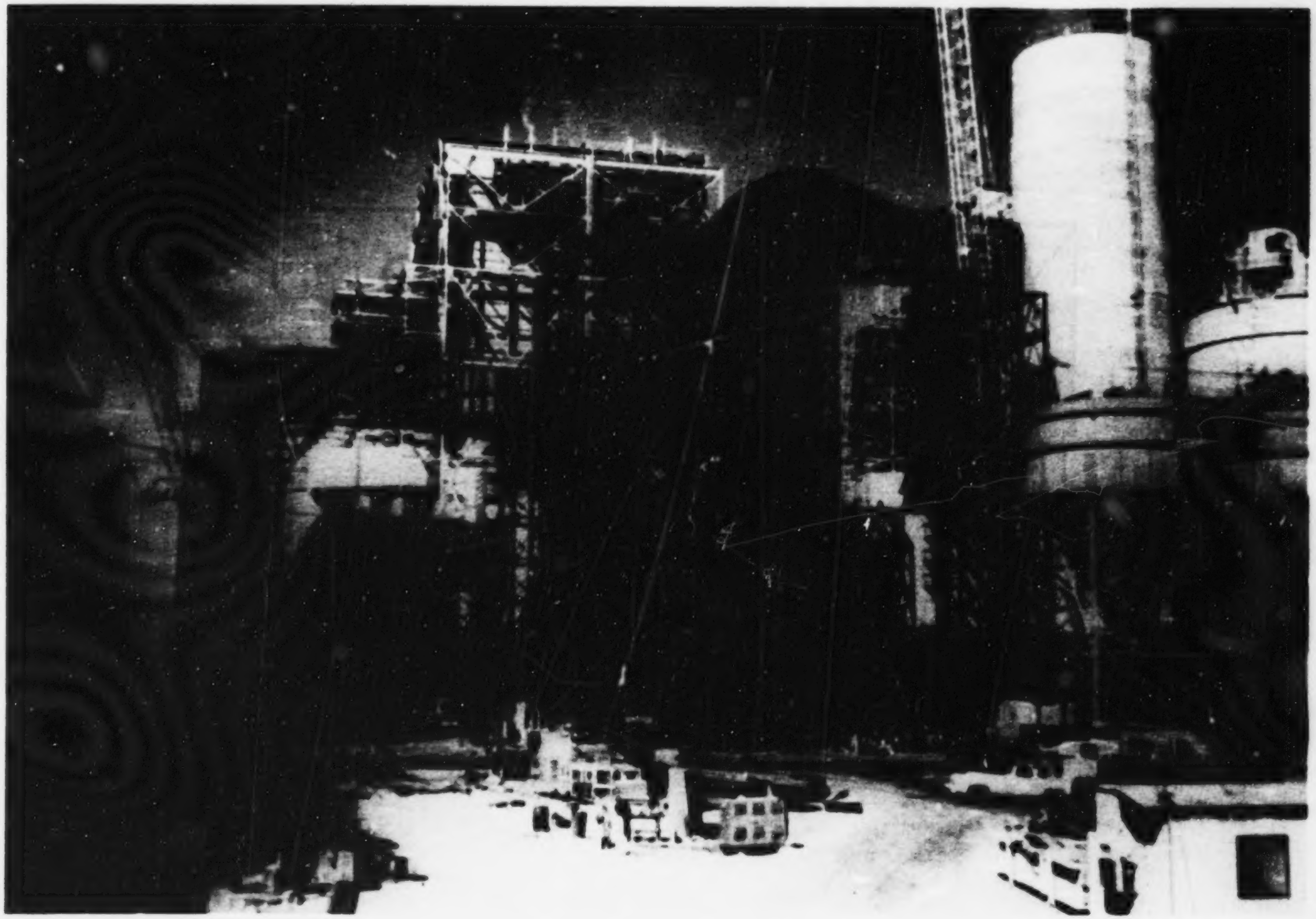


FIGURE 1 - 170 MW VERTICAL MODULE

Figure 2

170 MW VERTICAL MODULE (4 STAGE TCA)

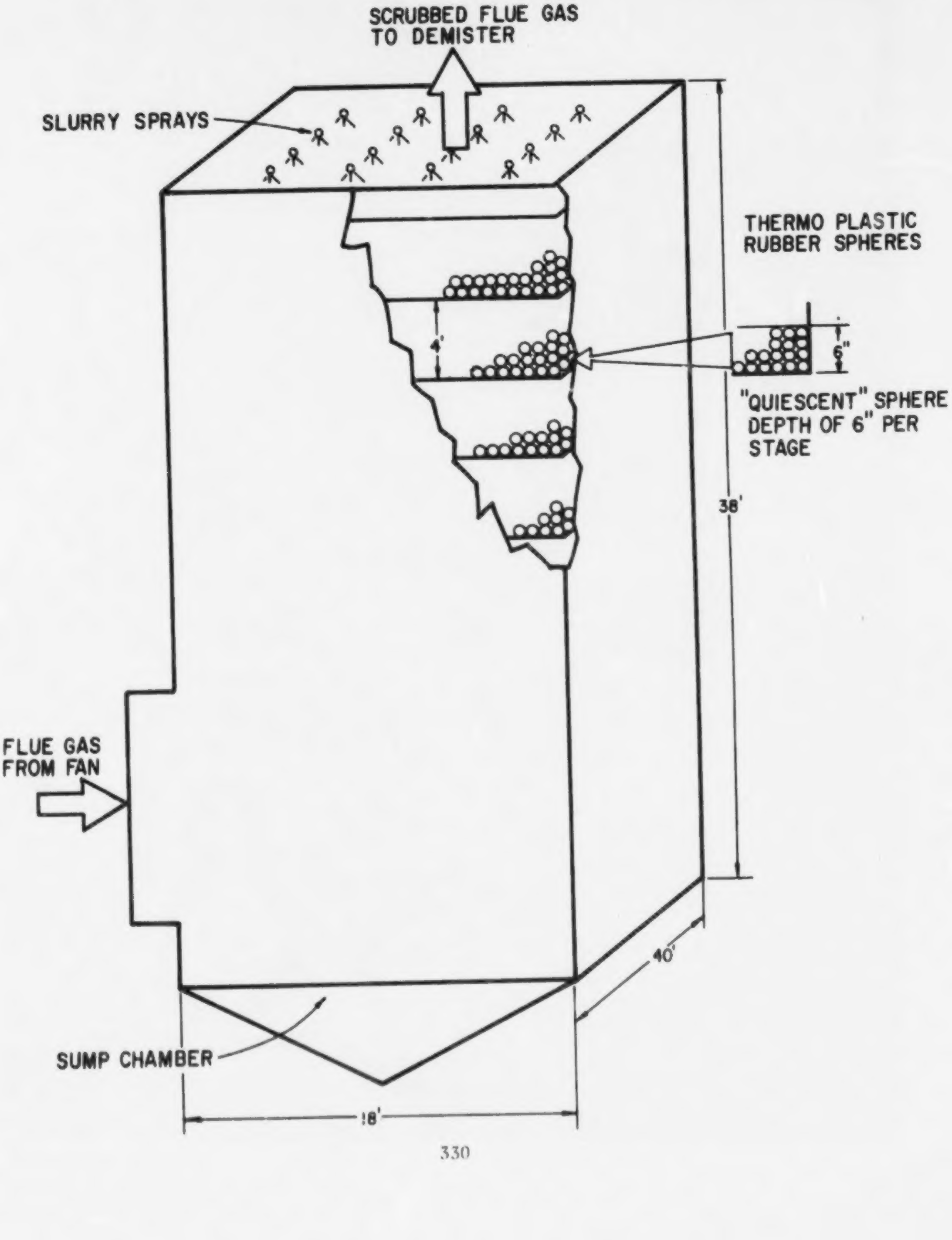
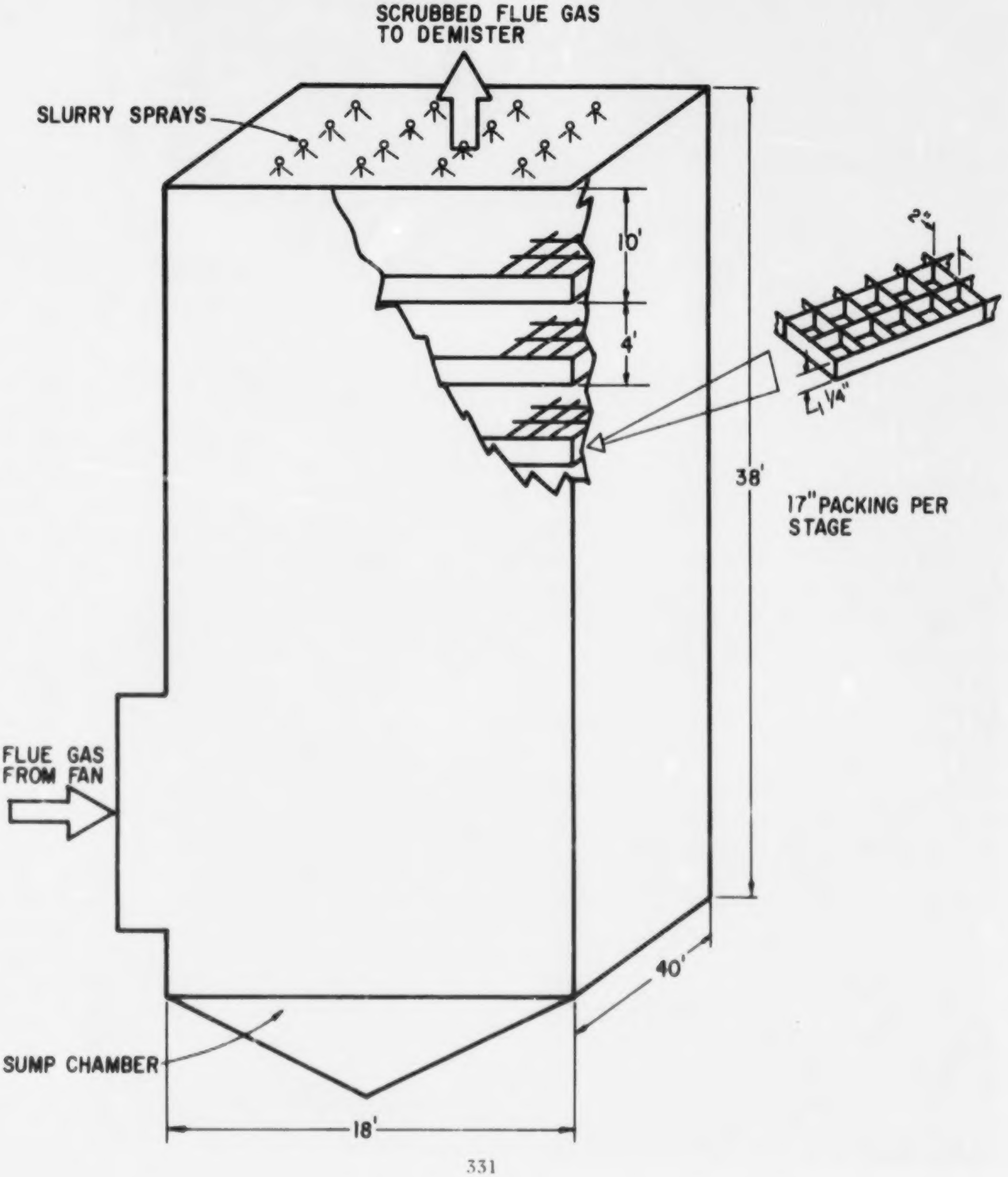

SCRUBBED FLUE GAS
TO DEMISTER
SLURRY SPRAYS
THERMO PLASTIC
RUBBER SPHERES
4'
6"
"QUIESCENT" SPHERE
DEPTH OF 6" PER
STAGE
38'
FLUE GAS
FROM FAN
40'
SUMP CHAMBER
18'

Figure 3

170MW VERTICAL MODULE
(3 STAGE POLYGRID PACKED ABSORBER)


SCRUBBED FLUE GAS
TO DEMISTER
SLURRY SPRAYS
10'
4'
2"
1 1/4"
38'
17" PACKING PER
STAGE
FLUE GAS
FROM FAN
40'
SUMP CHAMBER
18'

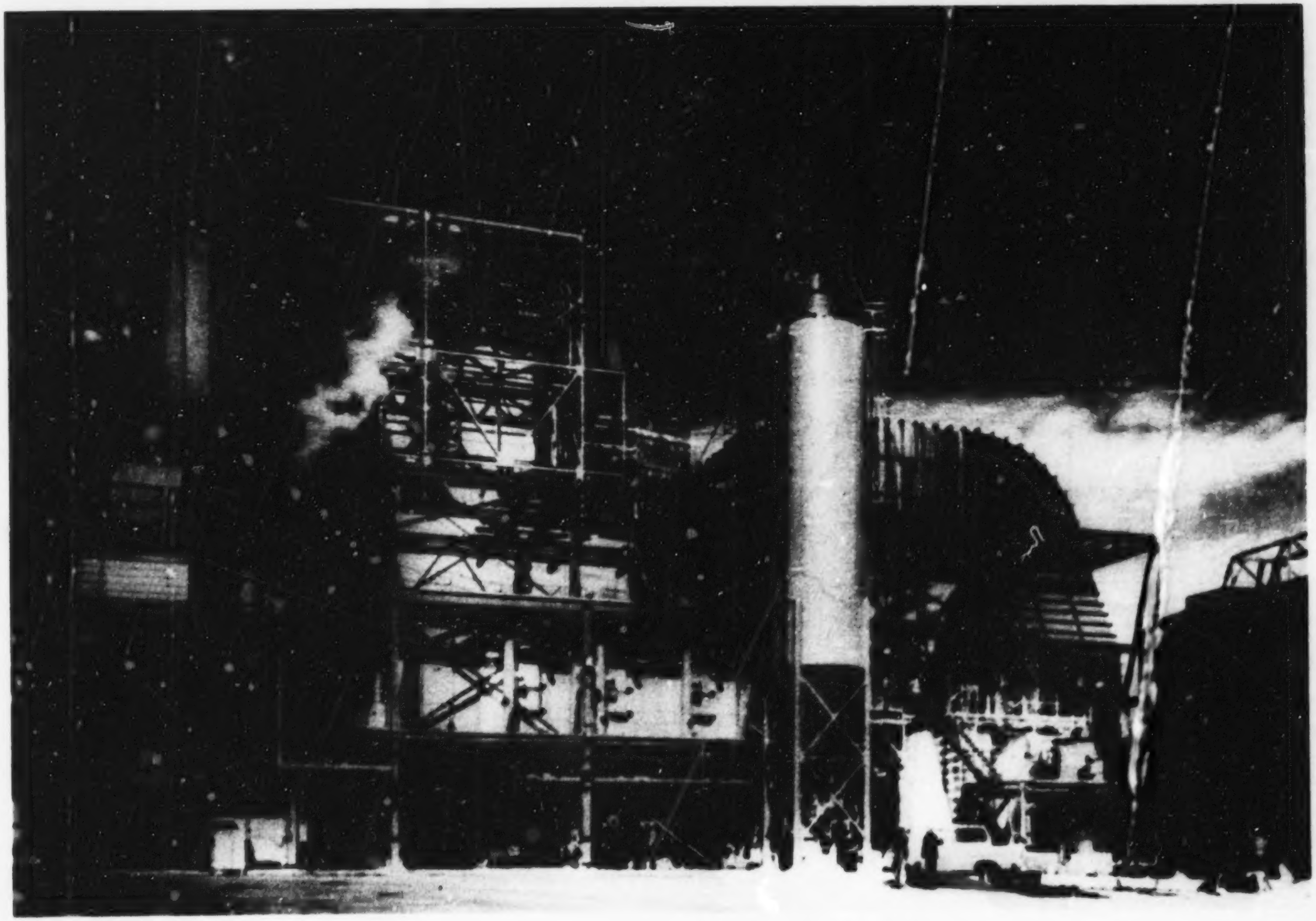


FIGURE 4 - 170 MW HORIZONTAL MODULE

4 supplies flue gas to the scrubber from the downstream side of the Unit 2 electrostatic precipitator. A 1,750 HP booster fan forces the flue gas through the scrubber shell and demister section into the outlet ducting where it is heated by injecting hot ambient air and is finally returned to the precipitator outlet duct upstream from the 500 foot stack.

The four-stage scrubbing chamber of the Horizontal Module is shown in Figure 5 and is 48 feet long, 28 feet wide and 15 feet high. The Horizontal Module did not contain packing, but consisted of four stages of cross flow spray. The slurry was cycled through the scrubber in a countercurrent manner. That is, the fresh lime slurry from the mix tank was first sprayed across the flue gas at the fourth stage, or discharge end of the scrubbing chamber. The same liquid was successively collected and pumped to the third, second, and first stages and successively depleted of alkalinity. By the time the slurry reached the first stage collection hopper for return to the lime mix tank, it was almost completely depleted of any excess alkalinity.

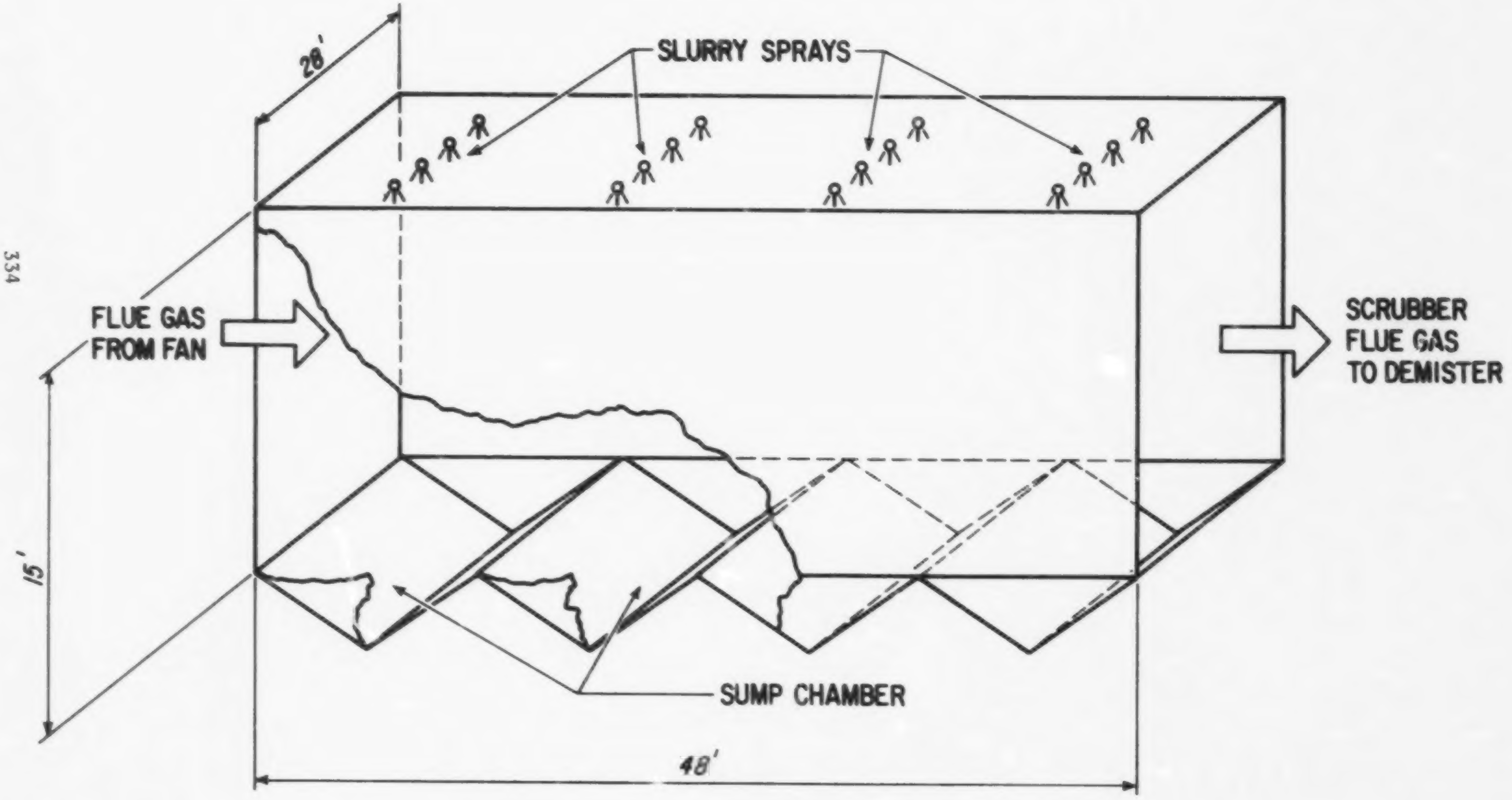
The sprays were discharged from a row of 36 externally-mounted nozzles at each stage. The recirculating slurry flow rate was normally 9,000 gpm, but the slurry was mechanically pumped four times per circuit for a total installed pump capacity of 36,000 gpm. As in the Vertical Module, a small quantity of recirculating slurry was discharged to a dewatering complex to extract solids from the scrubbing system. Water reclaimed in the dewatering process was returned to the scrubber slurry lime mix tank. However, in the case of the Horizontal Module, the dewatering complex consisted of a thickener tank and a sludge disposal pond.

TEST CONDITIONS

Both test modules were tested under a variety of configurations and conditions to determine the significant parameters affecting SO2 and particulate removal, operating simplicity and system chemistry. The test program for each module was divided into test blocks and were designed with specific values for selected variables. This approach was adopted to simplify data analysis and permit the effect of each variable to be determined independently. The variables considered most relevant are listed below:

Type of reagent
Demister wash procedure
Number of scrubbing stages
Percent cooling tower blowdown in makeup water
Spray nozzle pressure
Flue gas flow rate
Circulating liquid flow rate
Inlet SO2 concentration
Inlet particulate grain loading
pH of the scrubber slurry
Method of water balance control
Percent solids in scrubber slurry
Percent solids in reagent feed slurry

FIGURE 5
170 MW HORIZONTAL MODULE
(4 STAGE)


SLURRY SPRAYS
28'
FLUE GAS FROM FAN
SCRUBBER FLUE GAS TO DEMISTER
15'
SUMP CHAMBER
48'

TEST PERFORMANCE RESULTS

SO2 Removal

A comparison of SO2 removal efficiency as a function of recirculating slurry flow rate for the Horizontal and Vertical Modules is shown in Figure 6. Note that with low sulfur Western coal fired at the Mohave Generating Station, the SO2 concentration in the stack gas and at the scrubber inlet averaged only 200 ppm. At this low inlet SO2 concentration, SO2 removal above 95% was achieved with all three scrubbers. For example, the recirculating liquid flow rate required for 96% SO2 removal was 9,500 gpm with the Horizontal Module 18,500 gpm with the Vertical Module in the TCA mode and 27,800 gpm with the Vertical Module in the PPA mode.

Effect of Staging on SO2 Removal

The percentage SO2 removal in the Horizontal Module as a function of the number of scrubbing stages is shown in Figure 7. The first stage removes 50% of the SO2. The second stage removes 64% of the remaining SO2 for a total SO2 removal with two stages of 82%. Each successive stage removes some of the remaining SO2 and although the total percentage removal increases, the total weight of SO2 removed in each additional stage decreases. The Horizontal Module design allows a selection of the appropriate number of scrubbing stages to obtain the desired degree of SO2 removal. Since the Horizontal Module can continue to operate with the loss of all but one stage, it was relatively easy to obtain the effect of staging by selectively shutting off stages. It was more difficult to obtain this type of data with the Vertical Module since shutdown was required to remove the packing. However, the Vertical Module was operated in the TCA mode with 3 and 4 stages and in the PPA mode with 2 and 3 stages. The results of the effect of staging on SO2 removal were in general agreement with those obtained with the Horizontal Module.

Effect of Turndown on SO2 Removal

All three scrubbers showed a slightly increased degree of SO2 removal when the gas flow rate was decreased from 450,000 SCFM and the circulating liquor flow rate was allowed to remain constant. It should be recognized that the SO2 concentrations at the exit of the scrubber were extremely low and wet chemical analysis, rather than instrument readings, were used. The measured exit valves were corrected (i.e. increased) for the dilution effects of water vapor, and in the case of the Horizontal Module for the reheat air so that they were comparable to the inlet values. The effect of turndown on exit SO2 concentration is shown in tabular form in Figure 8.

Particulate Removal and Turndown

At the design operating condition of 450,000 SCFM and 0.10 gr/SCF inlet particulate grain loading, both the Horizontal and Vertical Modules achieved a relatively high degree of particulate removal. This is shown in Figure 9, where the Vertical TCA achieved 93% removal, the Horizontal Module ahieved 92.5% removal and the Vertical PPA

FIGURE 6

EFFECT OF CIRCULATING LIQUOR FLOW RATE ON SO_2 REMOVAL AT CONSTANT GAS FLOW

G=450,000 SCFM

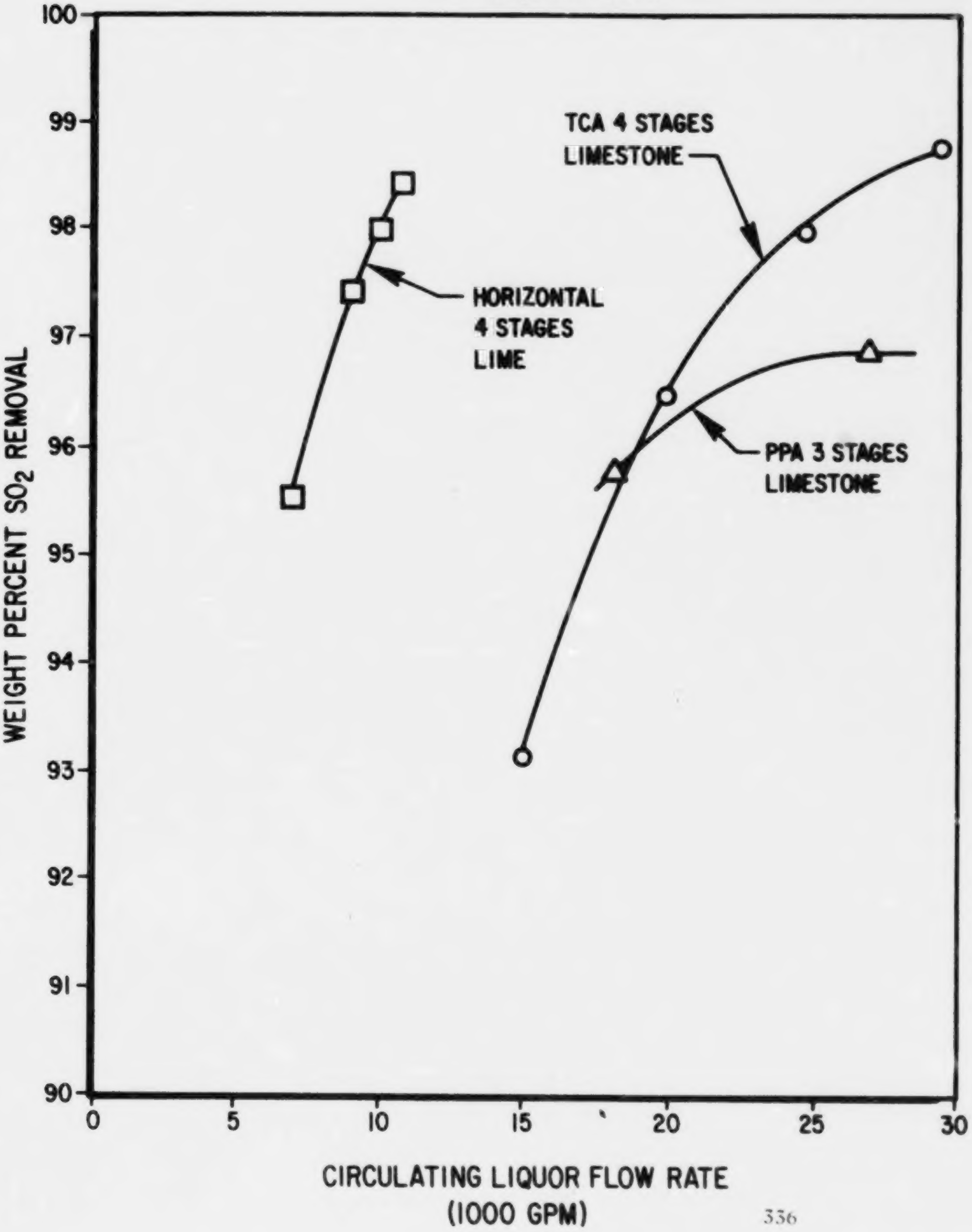


FIGURE 7
170 MW HORIZONTAL
SO_2 REMOVAL VS. NUMBER OF STAGES

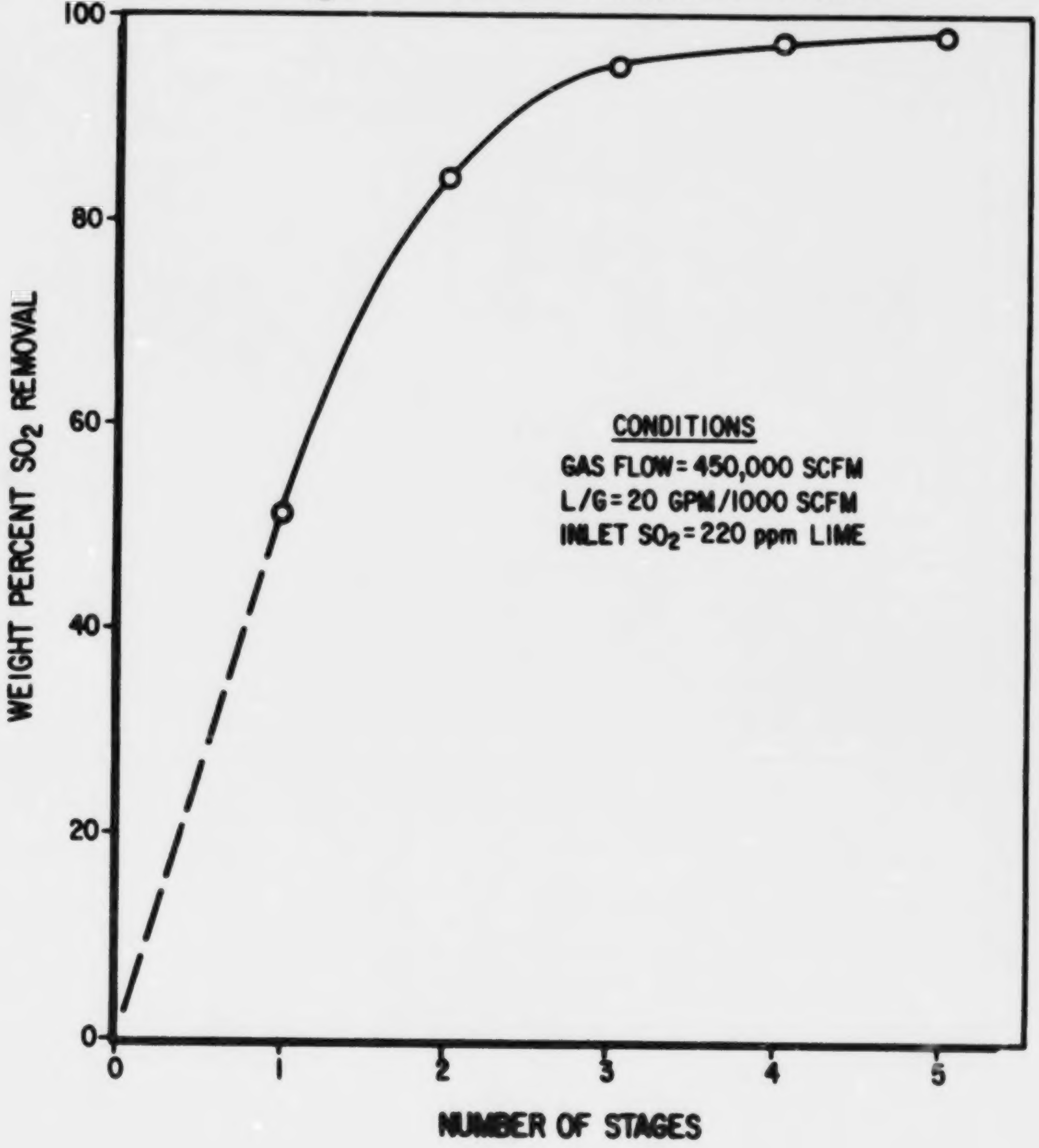
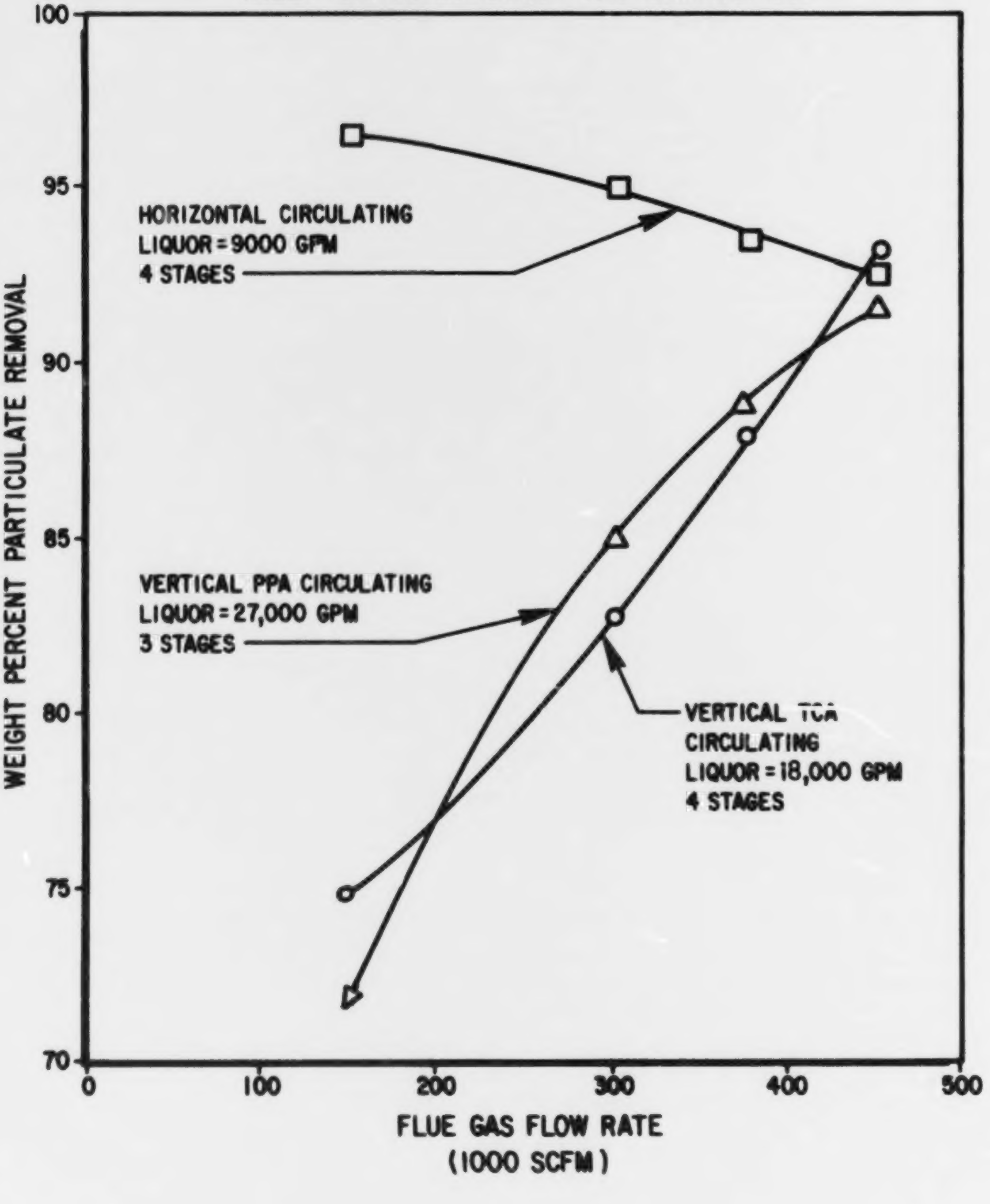


FIGURE 8

EFFECT OF TURNDOWN ON SO_2 REMOVAL

	HORIZONTAL	VERTICAL TCA	VERTICAL PPA
NUMBER OF STAGES	4	4	3
REAGENT	LIME	LIMESTONE	LIMESTONE
CIRCULATING LIQUOR FLOW RATE (GPM)	9,500	18,000	27,000
INLET SO_2 CONCENTRATION (ppm)	220	220	220
EXIT SO_2 CONCENTRATION (ppm)			
at 450,000 SCFM INLET	4	11	8
at 300,000 SCFM INLET	2	9	3

FIGURE 9
EFFECT OF TURNDOWN ON PARTICULATE REMOVAL
INLET GRAIN LOADING = 0.10 GRAINS/SCF



achieved 91.5% removal. Note that this is a relatively high weight percentage removal for the fine particles remaining in the flue gas after passing through an upstream electrostatic precipitator.

Note that the particulate removal efficiency of the Vertical TCA and PPA decreases as the flue gas flow rate is reduced, while the particulate removal efficiency of the Horizontal Module increases as a function of turndown ratio.

The Vertical TCA and PPA Modules achieve increased particulate removal at increased flue gas pressure drop levels. For example, at 450,000 SCFM, the pressure drop across the scrubbing chamber was 14 inches of water for the TCA and 12.2 inches for the PPA. At 33% of rated capacity with the design liquid flow rate the pressure drop was 3.5 inches of water in the TCA and 2.0 inches in the PPA. Note that at one-third of rated capacity with the design liquid flow rate, the liquid to gas contacting ratios are three times higher than the design value. However, the percentage particulate removal decreased with decreasing flue gas pressure drop from 93% to about 70%.

In the case of the Horizontal Module, particulate removal is primarily a function of the volume of flue gas contacted by the falling spray droplets. As flue gas flow rate is turned down from 450,000 SCFM to 150,000 SCFM, the pressure drop across the scrubbing chamber decreases from 1.0 inches of water to 0.10 inches of water. Since the liquid flow rate is constant, a given amount of the gas is contacted by three times as many droplets at one-third load as at full load. The particulate removal efficiency was observed to increase from 92.5% at full load to 96.5% at one-third load.

Closed Loop Water System

One of the objectives of the Test Modules Program at Mohave was to demonstrate the operation of a closed loop water system. This means that all the liquid water discharged from the scrubber is recycled for use except the water which cannot be reclaimed from the waste sludge. It should be noted that two molecules of water for each molecule of SO_2 removed from the flue gas exist in the sulfate sludge as water of hydration, or 36 pounds of water for each 32 pounds of sulfur. Since wet scrubbers act to cool and saturate hot inlet flue gas, a majority of the makeup water requirement in a closed loop scrubbing system is to replace the water discharged as water vapor from the stack.

A second objective of the Test Modules Program was to utilize up to 75% cooling tower blowdown water in the makeup water for a closed loop scrubbing system. The cooling tower blowdown water contained approximately 12,000 ppm total dissolved solids. The water balances measured during the test period are shown in Figure 10. The Horizontal Module was capable of operating with 75% cooling tower blowdown in the makeup and the Vertical PPA was capable of operating with 40% blowdown in the makeup for scale-free operation during extended testing.[a] Prior to converting the Vertical Module to the PPA mode, considerable scaling occurred when the Vertical Module was operated with 65% cooling tower blowdown in the makeup water. This scaling may have been due to the presaturator configuration, later changed, rather than solely due to tha amount of cooling tower blowdown water used as make up. This prevented obtaining satisfactory water balance data as is shown in Figure 10 for the other two scrubbers.

---

[a] The two stage PPA configuration was operated for 502 hours with 83% cooling tower blowdown water as makeup.

Both modules thus demonstrated the ability to achieve over 90% of the makeup water discharged from the stack as pure water vapor. Note that the Horizontal Module was capable of scale-free operation at a slurry dissolved solids level of 185,000 ppm. The Vertical PPA slurry dissolved solids level was 70,000 ppm at the reduced consumption of cooling tower blowdown required for scale-free operation. The dissolved salts represented were largely sodium chloride (NaCl), magnesium sulfate (MgSO4) and sodium sulfate (Na2 SO4).

Table 11-2 of the March 1975 National Academy report indicated that closed loop operation of the Horizontal Module was achieved at Mohave, but indicated in a footnote "Mohave and Cholla experience little rainfall and water losses due to evaporation from their sludge ponds are significant". It is true that Mohave experiences little rainfall, but it is not believed that a 4 gpm loss by evaporation from the sludge pond is significant compared to the 156 gpm makeup water requirement. The presence of rainfall had no measurable influence on the Vertical Module where centrifuging or filtering was used as the dewatering technique. The liquid water blowdown rate in the Vertical Module was 6.5 gpm compared to a 160 gpm makeup water rate.

The key to closed loop operation is in the dewatering complex, where water is reclaimed from purged scrubber solids and returned to the scrubbing system. This allows the makeup water flow rate to be a function primarily of flue gas flow rate by providing all internal water requirements with recycled water. The scrubbing system could thus be turned down to any desired capacity while simultaneously turning down the makeup water flow rate to avoid operating the system out of water balance. The most critical part of the water recycle concept required for closed loop operation was the source of pump and instrument seal water. The seal water requirement for each module represented a flow rate of about 50 gpm, which was independent of flue gas flow rate, SO2 content or unit load. For both Horizontal and Vertical scrubbing systems, thickener overflow was filtered to provide the pump and instrument seal water. In spite of the high dissolved solids in the seal water no severe chemical scaling or corrosion was observed.

Operating Conditions at Design Gas Flow Rate

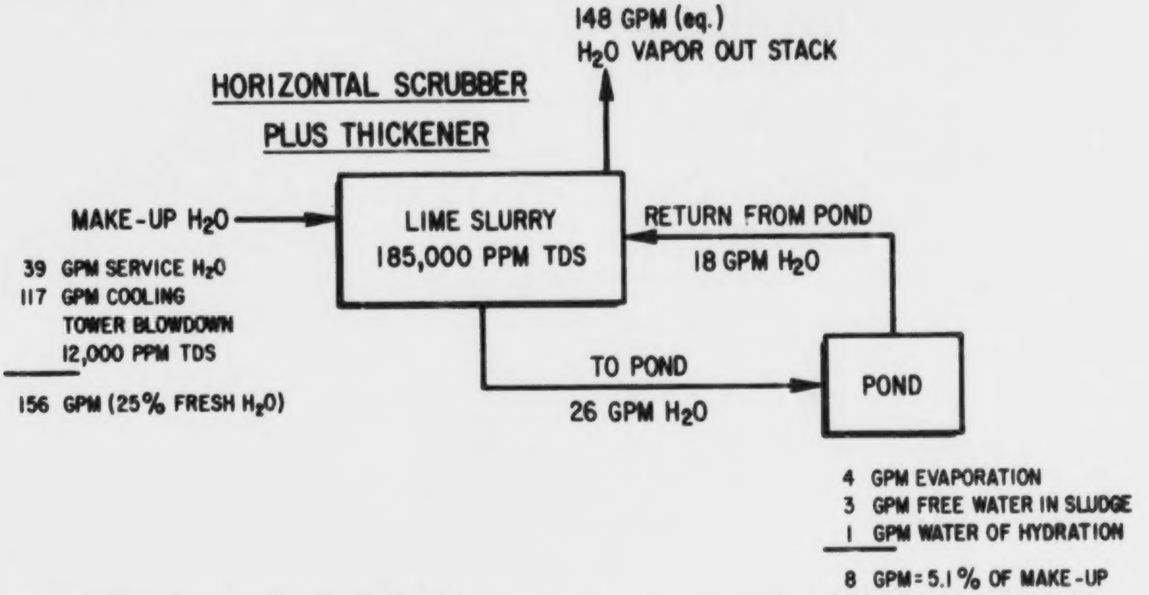
The observed operating conditions at the design gas flow rate of 450,000 SCFM are summarized for the three types of scrubbers in Figure 11. Several items deserve comment.

Electric Power Requirements

The flue gas pressure drop across the scrubbing chamber was an order of magnitude less for the Horizontal Module than for the Vertical TCA or PPA. The electric power requirements for the Vertical Module

FIGURE 10
WATER BALANCES

INLET GAS TEMPERATURE = 270° F.
450,000 SCFM


HORIZONTAL SCRUBBER PLUS THICKENER
148 GPM (eq.) H_2O VAPOR OUT STACK
MAKE-UP H_2O
39 GPM SERVICE H_2O
117 GPM COOLING TOWER BLOWDOWN 12,000 PPM TDS
156 GPM (25% FRESH H_2O)
LIME SLURRY 185,000 PPM TDS
RETURN FROM POND
18 GPM H_2O
TO POND
26 GPM H_2O
POND
4 GPM EVAPORATION
3 GPM FREE WATER IN SLUDGE
1 GPM WATER OF HYDRATION
8 GPM = 5.1% OF MAKE-UP

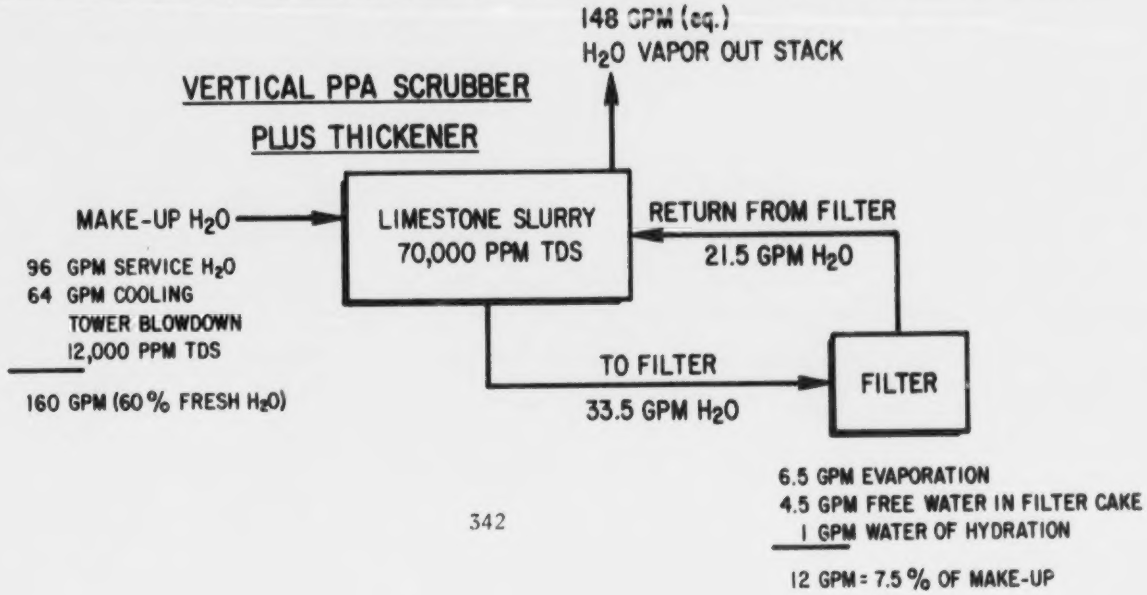

VERTICAL PPA SCRUBBER PLUS THICKENER
148 GPM (eq.) H_2O VAPOR OUT STACK
MAKE-UP H_2O
96 GPM SERVICE H_2O
64 GPM COOLING TOWER BLOWDOWN 12,000 PPM TDS
160 GPM (60% FRESH H_2O)
LIMESTONE SLURRY 70,000 PPM TDS
RETURN FROM FILTER
21.5 GPM H_2O
TO FILTER
33.5 GPM H_2O
FILTER
6.5 GPM EVAPORATION
4.5 GPM FREE WATER IN FILTER CAKE
1 GPM WATER OF HYDRATION
12 GPM = 7.5% OF MAKE-UP

FIGURE II
OPERATING CONDITIONS AT DESIGN GAS FLOWRATE
450,000 SCFM

	HORIZONTAL (4 STAGES)	VERTICAL TCA (4 STAGES)	VERTICAL PPA (3 STAGES)
L/G RATIO (GPM/1000 SCF)	21	36	60
SCRUBBER PRESSURE DROP (IN H_2O)	1.0	14.0	12.2
SYSTEM ELECTRIC POWER REQUIRED (MW)	2.6	3.4	3.9
REAGENT TYPE	LIME	LIMESTONE	LIMESTONE
SPENT SLURRY pH	6.2	5.4	5.1
REAGENT UTILIZATION	99%	75%	92%
PERCENT COOLING TOWER BLOWDOWN H_2O	75%	65%	41%
PERCENT SLURRY SOLIDS	5%	5%	15%
FORMATION OF GYPSUM SCALE	NO	YES (ESPECIALLY ON FIRST STAGE GRIDS)	NO

were consequently 30% to 50% higher than for the Horizontal Module. For example, at design operating Conditions, the measured electric power consumption of the Vertical Module was 3.4 MW in the TCA mode and 3.9 MW in the PPA mode, while the power consumption of the Horizontal Module was 2.6 MW. This included all of the power requirement including air conditioning of the control rooms.

Reagent Comparison

Figure 11 indicates a lime utilization of 99% in the Horizontal Scrubber. The methods used to measure this were reported in a previous paper.[2] Figure 11 also indicates a limestone utilization of 65% with 5% slurry solids in the TCA scrubber and 92% utilization with 15% slurry solids in the PPA scrubber. The role of percent slurry solids in influencing limestone utilization was also presented previously.[2] Increased slurry solids increase limestone utilization but also increase wear on pumps and nozzles as well as increase the possibility of solids buildup in the scrubbing system components. Both the lime and limestone utilization data obtained in the 170 MW scrubbers compare well with the data obtained on the pilot plant scale. Utilizing 95% pure reagents, the data in Figure 11 can be used to calculate that 2.75 times as much limestone by weight as lime is required to remove a given amount of SO2.

The choice between lime and limestone is a trade off between a number of factors in which the quantity of reagent, including the associated transportation cost, is only one factor. The operating and capital costs of the equipment required to pulverize limestone to a 225 mesh size must be balanced against the costs of a lime slaker to slake 1/4 inch pebble lime. The decreased solubility of limestone, compared to lime, results in increased holding tank sizes unless special attention is paid to the chemistry of the scrubbing solution. Our experiments, both on the 1/2-1 MW size and the 170 MW size indicated that a higher L/G was required with limestone than with lime to achieve the same degree of SO2 removal, all other factors being equal. Finally, the unreacted limestone increases the amount of sludge which must be disposed of and it's presence makes more difficult the conversion of the sludge into potentially useful products such as gypsum wallboard. All in all, the cost per ton of a limestone versus lime is not a true measure of the actual costs which will be incurred.

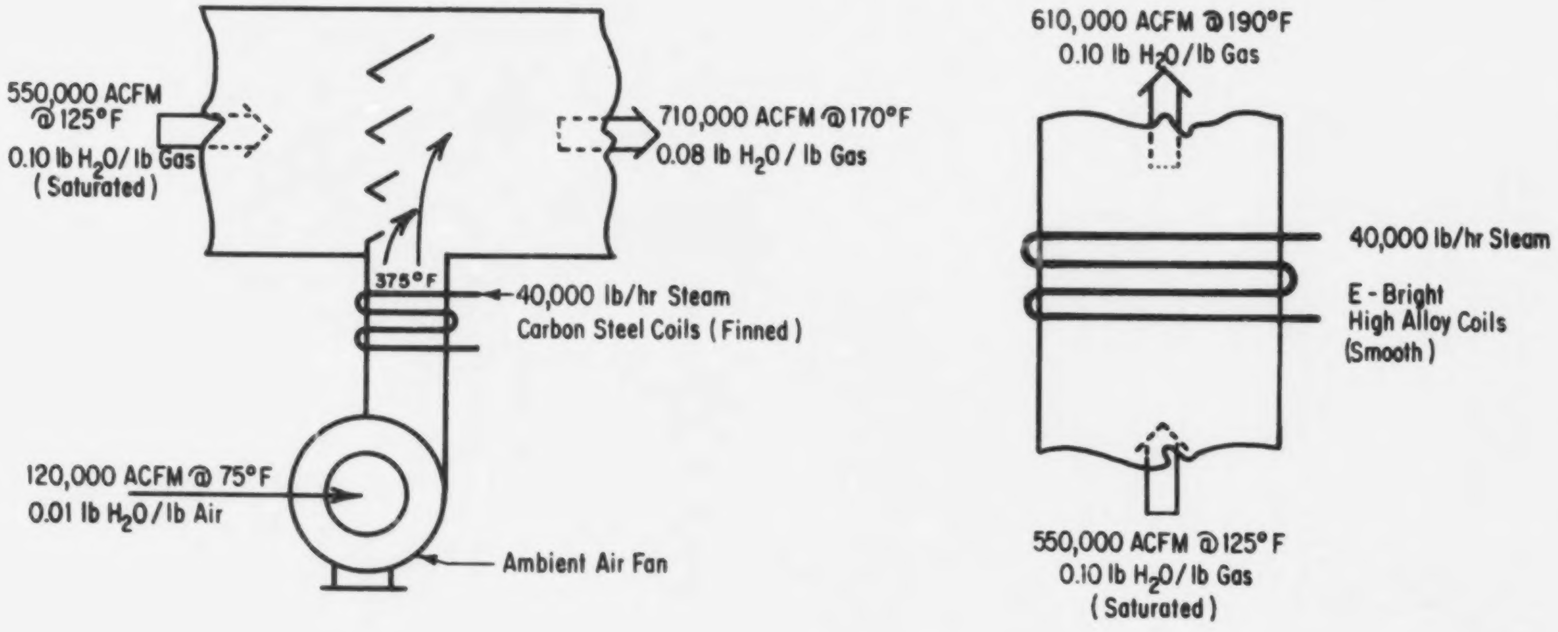
Reheaters

It is well known that wet scrubbers cool and saturate flue gas. Without flue gas reheat, the saturated scrubber exit gas would form a dense, white water vapor plume. On cold days this plume can persist for thousands of feet. Even with reheat, to 200 degrees Farenheit, prevention of a dense water vapor plume may be impossible on some winter mornings.

A comparison of the two methods of flue gas reheating tested at Mohave is shown in Figure 12. On the left is the indirect reheat method, where outside air is heated to about 375 degrees Farenheit with steam coils and introduced into the cold flue gas at the outlet of the scrubber. On the right is the direct reheat method, where steam coils in direct

Figure 12

REHEATER COMPARISON FOR EQUIVALENT REDUCTION IN FOG FORMATION



INDIRECT REHEAT

DIRECT REHEAT

550,000 ACFM
@125°F
0.10 lb H_2O/lb Gas
(Saturated)
710,000 ACFM @170°F
0.08 lb H_2O/lb Gas
375°F
40,000 lb/hr Steam
Carbon Steel Coils (Finned)
120,000 ACFM @ 75°F
0.01 lb H_2O/lb Air
Ambient Air Fan
610,000 ACFM @190°F
0.10 lb H_2O/lb Gas
40,000 lb/hr Steam
E - Bright
High Alloy Coils
(Smooth)
550,000 ACFM @125°F
0.10 lb H_2O/lb Gas
(Saturated)

contact with outlet flue gas provide the desired amount of reheat. (The so-called direct reheat method where oil or gas is burned directly in the exit flue gas was not tested in this program.) Both methods of reheat require the same amount of steam for the same degree of reduction in water vapor plume formation. Indirect reheat using relatively dry ambient air requires an exit temperature of only 170 degrees Farenheit to provide the same reduction in fog formation as 190 degrees Farenheit exit temperature with direct reheat. This is because the exit gas in the indirect reheat case is 20% diluted with relatively dry ambient air. However, plume buoyancy or possible stack condensation problems might dictate using a higher exit temperature gas.

During the test program, it was found that the ambient air fan used for indirect reheat provided a built-in means of positive pressure air purge when the scrubber was shut down. Guillotine dampers at both ends of the scrubber were under positive pressure with ambient air due to the indirect reheat fan, which allowed safe internal access to the ductwork without expensive pressure sealed double guillotine dampers. The indirect reheat steam coils were not subject to the same environment as the direct reheat coils. These findings led to the conclusion that indirect reheat was preferred compared with direct reheat.

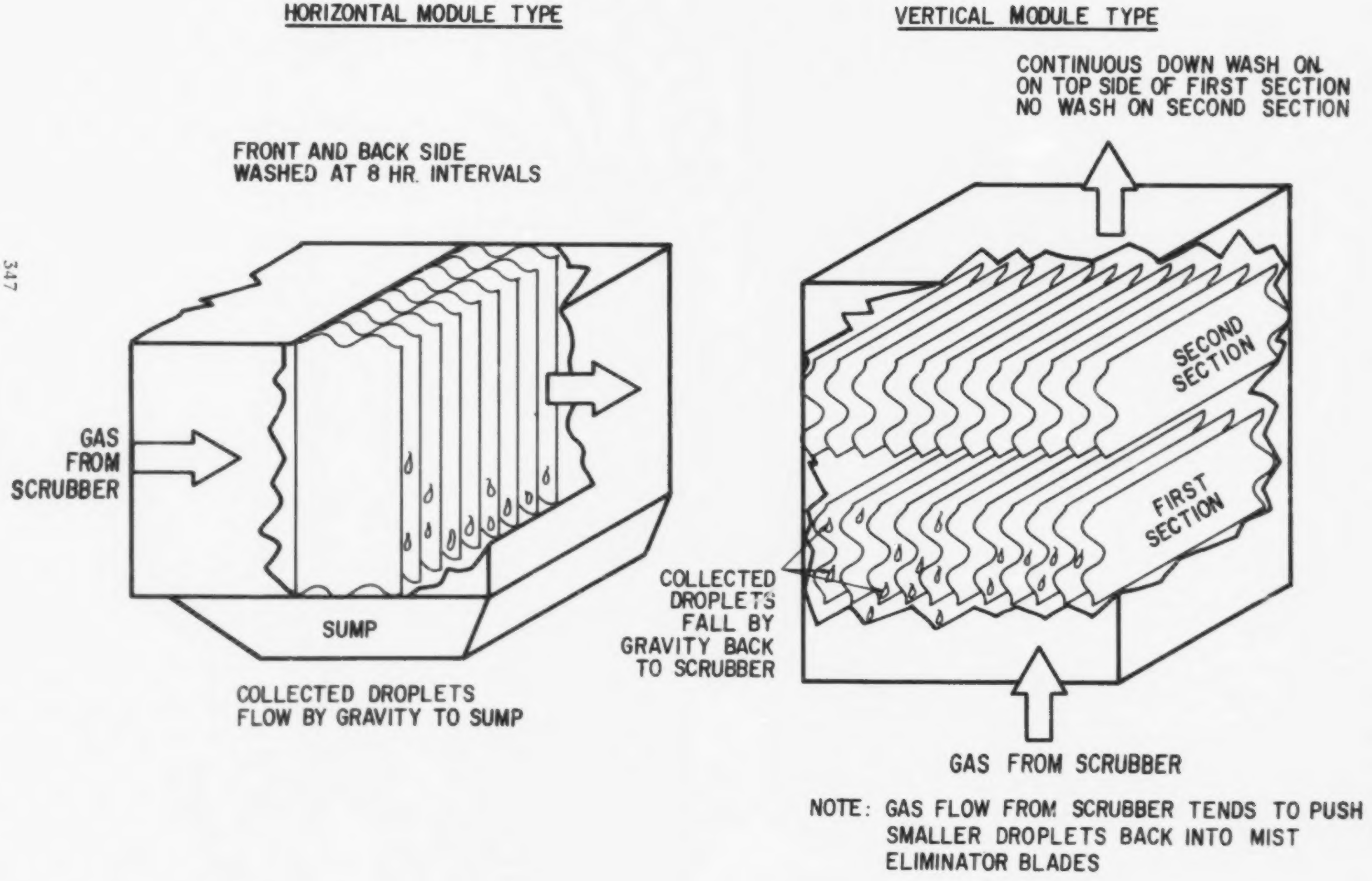
Mist Eliminators

Two types of mist eliminators were tested at Mohave, as shown in Figure 13. The horizontal type (gas flow in horizontal plane) utilized one mist eliminator section with front and back sides deluge washed at intermittant intervals. The normal wash cycle was about 15 minutes every eight hours. The vertical type utilized two mist eliminator sections with no wash on the second section and continuous downwash on the top side of the first section. Difficulty was experienced in keeping this mist eliminator wash pump operating continuously during the test program. Both mist eliminators used 3-pass blades as shown, with plastic blades in the horizontal type and stainless steel blades in the vertical type. The plastic blades were not capable of withstanding temperature over about 180 degrees Farenheit without warping. The vertical type of mist eliminator suffered the disadvantage of no capability for removing deposits from the second mist eliminator section without a scrubber shutdown for maintenance. Both types of mist eliminators operated satisfactorily during the test program, but there were deposits on the second section of the vertical-type mist eliminator blades that continued to build up. It was concluded that a horizontal-type mist eliminator with intermittant wash on the front and back sides was preferred to the two section vertical-type mist eliminator with the described wash configuration.

Unavailability - History During the Test Program

Unavailability is defined as the total time that the scrubber was shut down and could not be operated due to design deficiency or maintenance problems. Unavailability does not depend on whether or not the generating unit is in operation, while availability is frequently defined as a function of generating unit operation. In an installation which involves a number of modules connected to one generating unit,

Figure 13
MIST ELIMINATORS


HORIZONTAL MODULE TYPE

VERTICAL MODULE TYPE

FRONT AND BACK SIDE WASHED AT 8 HR. INTERVALS
GAS FROM SCRUBBER
SUMP
COLLECTED DROPLETS FLOW BY GRAVITY TO SUMP
CONTINUOUS DOWN WASH ON ON TOP SIDE OF FIRST SECTION NO WASH ON SECOND SECTION
SECOND SECTION
FIRST SECTION
COLLECTED DROPLETS FALL BY GRAVITY BACK TO SCRUBBER
GAS FROM SCRUBBER

NOTE: GAS FLOW FROM SCRUBBER TENDS TO PUSH SMALLER DROPLETS BACK INTO MIST ELIMINATOR BLADES the unavailability record is the best indicator of how much of the time a given module will not be available for service.

The unavailability history for the test modules program is given in Figure 14. As can be seen, for 100% availability from a series of modules, more modules would probably be required with the Vertical configuration than with the Horizontal configuration.

During the Test Program, design modifications were made to alleviate required maintenance and decrease downtime for maintenance. In the Horizontal Module, the mist eliminator blade wash system was improved and a single mist eliminator section was found to be satisfactory. Successful modifications were also made to the scrubber to alleviate inlet gas distribution problems. Methods were also successfully tested to repair worn pump impellers and change worn spray nozzles without shutting down the Horizontal Module. Spray nozzles made with refractory materials were found to have longer life (estimated over one year) than nozzles manufactured with other materials.

In the Vertical Module, many successful design modifications were also made. However, no methods were discovered for preventing migration of the thermoplastic rubber balls between compartments without shutting down the scrubber and redistributing the balls. Replacement or cleaning of slurry distribution nozzles in the Vertical Module because of plugging or wear also required shutting down the scrubber. Finally, scale formation at the bottom of the first stage grids was aggravated by gas flow distribution problems in the Vertical Module. The poor gas flow distribution at the first stage was never resolved, but is suspected to have aggravated ball migration and gas flow channeling problems in the Vertical Module.

DISCUSSION

There were several findings in this program which are contrary to commonly accepted beliefs. At the beginning of the pilot plant program it was believed by many people that it would not be possible to reach exit concentrations of SO2 lower than 20 ppm (90% SO2 removal from an inlet concentration of 200 ppm inlet gas) because of the reduced partial pressure of SO2 as the concentration approaches zero. While the pilot plant results[2] refuted this, the Test Modules Program confirmed that it is possible to achieve an exit SO2 concentration of 2 ppm (99% SO2 removal) on a large scale.

Many people believe today that chemistry plays an overwhelming role in scrubber technology. While chemistry is important, for example the settling rate in the thickener tank and the quantity of sludge produced (sulfite crystals have 1/2 molecule of water associated with them compared to 2 molecules of water with the sulfate), one of the test program findings was that fluid mechanics also plays an important role, particularly in low maintenance operation. The major influences on scrubber design and operation are segregated as shown in Figure 15.

Finally, during the Test Modules Program, twenty-four variables, which influence visual plume opacity were identified.[5] Some of these have previously been identified (particle size, and grain loading)

FIGURE 14

UNAVAILABILITY HISTORY - MOHAVE 170 MW TEST MODULES PROGRAM

JANUARY 16, 1974 TO FEBRUARY 9, 1975

HORIZONTAL	HOURS
1. MODIFY AND REPAIR PLASTIC DEMISTER BLADES	503
2. CORRECT BOOSTER FAN BALANCE PROBLEMS	317
3. REPAIR CRITICAL PUMPS	256
4. REPLACE WORN-OUT SPRAY NOZZLES	238
5. MODIFY INLET GAS FLOW DISTRIBUTION	162
6. REPAIR HOPPER LEAKS	135
7. REMOVE HARDHAT FROM THICKENER	82
8. MODIFY SLAKING WATER TO PREVENT SCALE	45
9. CONDUCT INSPECTIONS FOR LONG RUNS	19
TOTAL	1757
TOTAL CALENDAR TIME	9328

PERCENT UNAVAILABILITY = 18.7

NOVEMBER 2, 1974 TO JULY 2, 1975

VERTICAL	HOURS
1. REPAIR GRIDS AND REDISTRIBUTE TCA BALLS	710
2. CLEAN SCALE FROM SCRUBBER INTERNALS	344
3. REPAIR OR REPLACE PLUGGED NOZZLES	153
4. REPAIR LEAKS IN TRAP-OUT TRAY	120
5. REPAIR/REALIGN PPA PACKING	85
6. CORRECT BOOSTER FAN TRIP PROBLEMS	72
7. CONDUCT INSPECTIONS FOR LONG RUNS	55
8. REMOVE HARDHAT FROM THICKENER	46
TOTAL	1585
TOTAL CALENDAR TIME	5813

PERCENT UNAVAILABILITY = 27.2

FIGURE 15

MAJOR INFLUENCING TECHNOLOGY
IN SO_2 SCRUBBING SYSTEMS

FLUID DYNAMICS

- FAN
- PRESATURATOR
- SCRUBBER DESIGN
- SCRUBBER PUMPS
- SCRUBBER PACKINGS
- SPRAY NOZZLES
- DEMISTER
- REHEATER
- DUCTING
- THICKENER
- VACUUM FILTER
- CENTRIFUGE
- PIPING AND VALVES
- EROSION
- INSTRUMENTATION/CONTROLS

CHEMISTRY

- REAGENT FEED SYSTEM
- REAGENT TYPE
- pH CONTROL SYSTEM
- SCRUBBER LININGS
- CORROSION
- SCALE PREVENTION

but the experimental data indicated that a major variable apparently heretofore unrecognized was the altitude of the sun. Figure 16 indicates the visual opacity reading which would be obtained by a "perfect" observer if the Mohave Generating Station were located in Florida or in Washington. With the absolute emissions of particulate matter remaining constant, the predicted variation in opacity from 10% to 90% as shown in Figure 16 is due solely to effects of the altitude of the sun. To date, we have not been able to convince the EPA that the sun rises in the East and sets in the West or that the days are longer in the summer than the winter. However, we have only been trying for a year and hope for success in eliminating visual opacity regulations in the future.

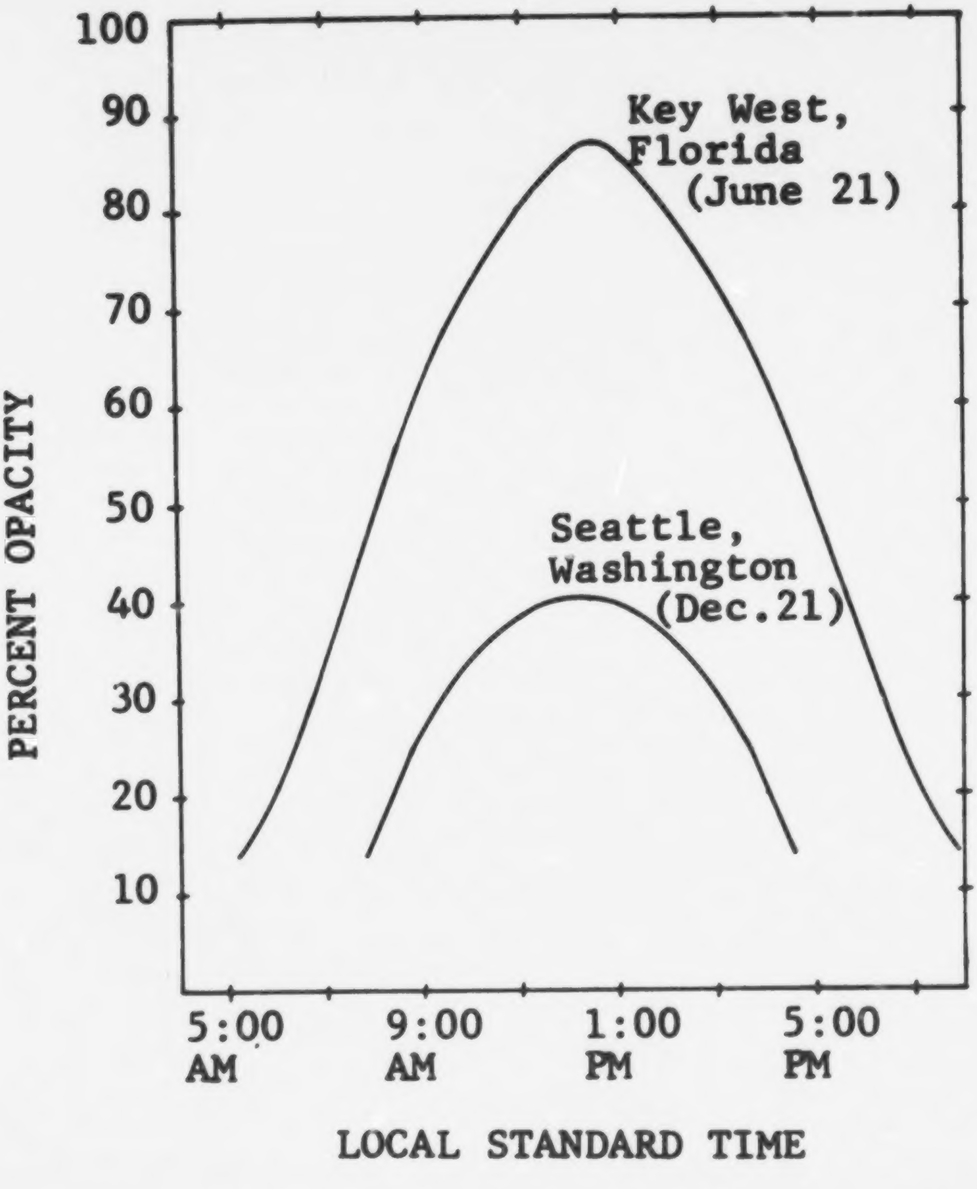


FIGURE 16 - EFFECT OF GEOGRAPHIC LOCATION AND TIME OF DAY ON OPACITY

BIBLIOGRAPHY

1. Shapiro,J.L. and Kuo,W.L. "The Mohave/Navajo Pilot Facility for Sulfur Dioxide REemoval" Second (2nd) EPA Flue Gas Desulfurization Symposium, November 8, 1971, Newe Orleans, Louisiana.

2. Weir,A., and Papay,L.T. "Scrubbing Experiments at the Mohave Generating Station" 3rd EPA Flue Gas Desulfurization symposium, May 14, 1973 - New Orleans, Louisiana.

3. Weir,A., Johnson,J.M., Jones,D.G., and Carlisle,S.T., "The Horizontal Crossflow Scrubber", 4th EPA Flue Gas Desulfurization Symposium, November 4, 1974, Atlanta, Georgia.

4. Handler,Philip et al "Air Quality and Stationary Source Control- a report by the commission on Natural Resources, National Academy of Sciences, National Academy of Engineering, and National Research Council - Prepared for the committee on Public Works, United States Senate - March, 1975 - Serial No. 94-4 U.S. Government Printing Office - Catalog Number Y4.P96/10:94-4

5. Weir,A., Jones,D.G., Papay,L.T., Calvert,S. and Yung,S. "Measurement of Particle Size and Other Factors influencing Plume Opacity" United Nations, U.S. EPA (and other) International Conference on Environmental Sensing and Assessment, September 14-19,1975, Las Vegas, Nevada.

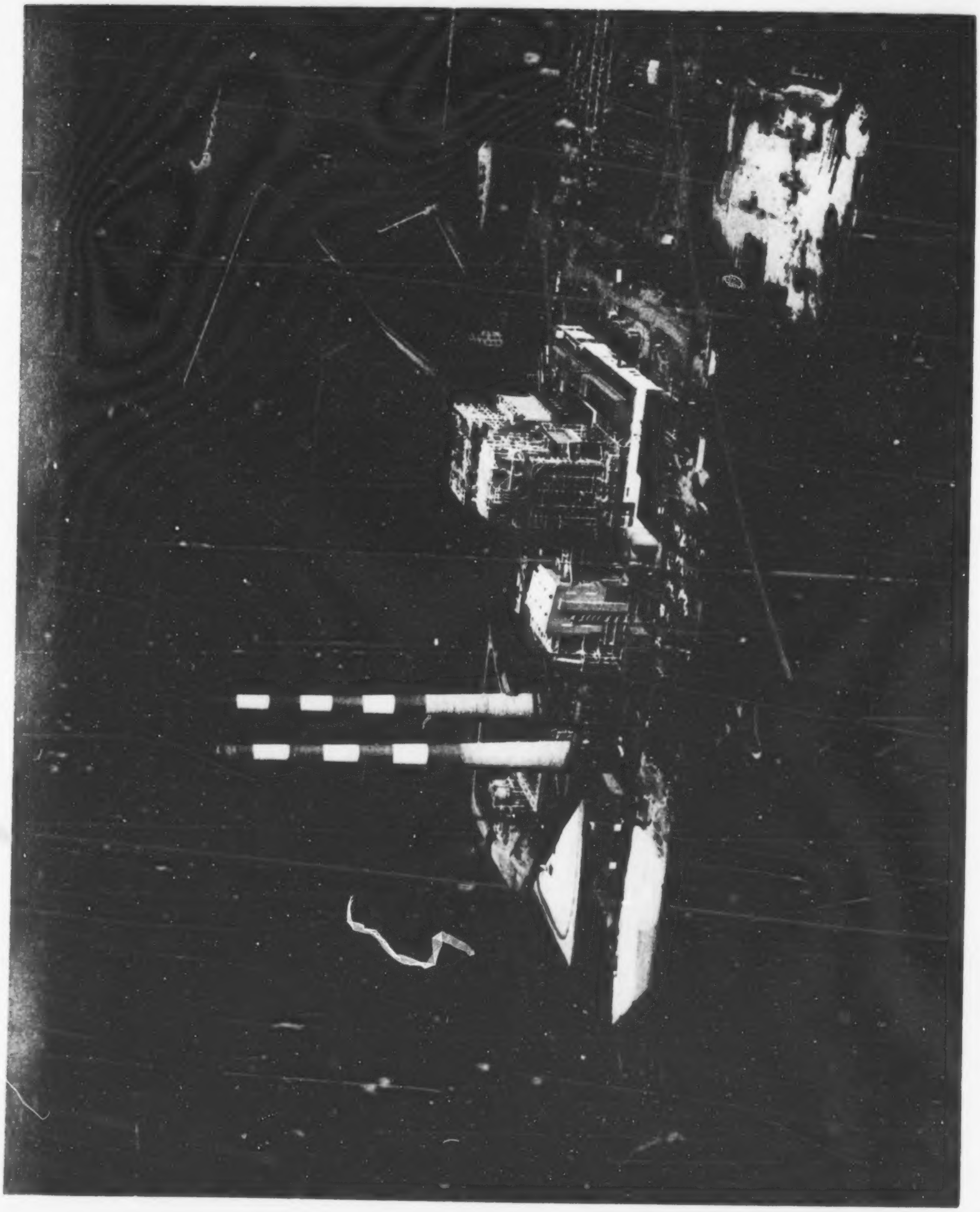
LA CYGNE STATION UNIT NO. 1
WET SCRUBBER OPERATING EXPERIENCE

Clifford F. McDaniel, Superintendent, Air Quality Control

Kansas City Power and Light Company
P. O. Box 211
La Cygne, Kansas 66040

ABSTRACT

This paper presents a description of the Babcock and Wilcox designed scrubbing system and reviews the status, costs, reliability and supportive data relating to our experiences in 1975.



LA CYGNE STATION UNIT NO. 1
WET SCRUBBER OPERATING EXPERIENCE

DESCRIPTION

The 820-megawatt La Cygne No. 1 unit began commercial operation on June 1, 1973, as a joint project of Kansas Gas and Electric Company and Kansas City Power & Light Company. The companies share equally in ownership and output and the unit is operated by KCPL. The 630-megawatt No. 2 unit, now about 45 per cent completed, is expected to be in service by the spring of 1977 under an identical arrangement.

The plant site is located about 55 miles south of downtown Kansas City, one-half mile west of the Missouri state line, and was selected based on locally available coal, water and limestone. Construction of No. 1 unit began in 1969 and erection of the Air Quality Control System was initiated in mid-1971.

Water for cooling purposes is furnished from a 2,600-acre reservoir constructed adjacent to the plant site. Fly ash and spent slurry from the AQC system is piped to a 160-acre settling pond located east of the reservoir.

Coal is delivered to the plant in off-the-road 120-ton trucks from surface mines operated by The Pittsburg & Midway Coal Mining Co. The nearby coal deposits are estimated to contain 70 million tons. The fuel is low grade, sub-bituminous with an as-fired heating value of 9,000 to 9,700 Btu/lb, and an ash content of 25 per cent and sulfur content of 5 per cent. (Exhibit A).

Limestone is obtained from nearby quarries and delivered to the plant in off-the-road 50-ton trucks.

The boiler for No. 1 unit is a cyclone-fired, supercritical, once-through, balanced-draft Babcock & Wilcox unit, with a rating of 6,200,000 pounds of steam per hour, 1,010 degrees F, 3,825 psig at the superheat outlet and 1,010 degrees F at the reheater outlet. The turbine-generator was supplied by Westinghouse and is rated at 874 MW gross output with five per cent overpressure and 3,500 psi throttle pressure. Three auxiliary, oil-fired boilers are used for plant start-up or for powering a 20 megawatt house turbine-generator. The net plant output is 820 megawatts, adjusted to include 24 megawatts used by the AQC system and 30 megawatts by plant auxiliaries.

PROCESS DESCRIPTION

The AQC system consists of seven identical two-stage Venturi-absorber scrubber modules (Exhibit B) designed to treat the boiler flue gas flow of 2,760,000 ACFM. (394,300 ACFM per module at 285 degrees F). The ductwork design does not provide for flue gas bypass of the system. Also, the plant does not have an alternate or secondary fuel supply. Each module can be isolated for maintenance by individual dampers. On site limestone grinding and slurry storage facilities provide up to 1,000 tons of slurry per hour. The unit has a balanced draft system with three 7,000 hp forced draft fans and six 7,000 hp induced draft fans located between the AQC system and the 700 foot stack. There is a common plenum at both the scrubber inlet and outlet. Spent slurry and fly ash are removed from the module recirculation tank through rubber lined pipes to the settling pond at the rate of 3,500 tons of solids per day. Clear make-up water is pumped from the pond and the loop is closed by recycling ball mill and module make-up water back into the system.

In abbreviated terms, as the hot flue gas enters the Venturi (Exhibit C), it is sprayed with slurry from 48 spray and 32 wall wash nozzles resulting in up to 99 per cent of the particulates agglomerated to the sump below. The gas continues through the sump making a 180 degree turn up through the absorber section. In the reaction chamber, the SO_2 is removed as the gas is forced through a limestone slurry solution sprayed on stainless steel sieve trays. The chemical reaction in part combines the calcium carbonate, water and sulfur dioxide to form two relatively insoluble calcium salts: calcium sulfate and calcium sulfite, which also fall to the sump. The cleaned gas passes through demisters to remove moisture and then is reheated to avoid deposits on the fans and provide a plume effect from the stack.

OPERATING EXPERIENCE

As a result of the continuing modification work, the availability of modules has been improved substantially, and for 1975, the total availability (Exhibit E) averaged 84.33 per cent, including both working and reserve hours.

The results of a stack sampling test conducted on May 5, 1975 (Exhibit H) with the unit continuously operating from 700 to 720 megawatts indicated an average sulfur dioxide removal efficiency of 80.14 per cent, and removal of 98.2 per cent of the particulates.

The ambient monitoring system indicates ground level concentrations well below the national standards for sulfur dioxide and nitrogen compounds (Exhibit I).

During a four hour, full load test in March (Exhibit F), the unit maintained a minimum of 800 megawatts with all seven modules fully loaded. The six induced draft fans were operating at less than maximum load and the three forced draft fans had not reached maximum rating. Sulfur dioxide removal efficiency for the six modules from which results were obtained was 76.2 per cent.

MAINTENANCE

In addition to the modifications, the improved performance of the AQC system has depended upon a substantial maintenance effort. Present procedures call for cleaning one module each night on a rotational schedule and keeping all modules available during the daytime peak periods. Cleaning requires three men from 10 to 12 hours, including time taken to open the many sections and place holds for personal safety. Recent modifications have greatly improved cleaning requirements and we are looking forward to modules staying on line continuously for up to three weeks.

Areas requiring attention are reheater pluggage; demister pluggage; Venturi wall and nozzle deposits, and sump accumulation. Hard scale has in the past been an enormous problem, especially in the sieve trays, but by closely controlling pH, severe scaling is not usually one of our chief troubles.

Carry-over to the induced draft fan blades continues to be one of our main operating concerns. We wash each individual fan about every four or five days. This requires a fan outage of from four to ten hours depending upon the necessity of a second or even a third washing. This is done with very high pressure water from 4,000 to 7,000 psi. The fan is taken off when the vibration from inbalance exceeds 12 Mils on either bearing.

To resist the effects of acid corrosion, various epoxy paints have been used, the most recent a Plasite #4030 black paint with a durakane resin base. This very tough surface has good erosion and corrosion resistance but eventually cracks and should be reapplied every 8 to 10 months to maintain wheel integrity. Currently we are considering protecting those areas most affected by acid runs by cladding with Inconel 625.

MANPOWER REQUIREMENTS

The scrubber operating and maintenance force currently totals 51 people and the organization is separate from the rest of the plant (Exhibit J). This is still considered to be a research and development situation and once routine operation is determined, the number of scrubber personnel will be reassessed.

COSTS

The total cost (Exhibit G) of the AQC system to date is $43 million, or about 22 per cent of the $200 million total plant cost, or about $52 a kilowatt installed. It is estimated that an additional $7 million investment will be required to reach optimum system performance. These figures do not include the allowance for funds used during construction. To date, the City of La Cygne, Kansas, has issued $30 million in tax exempt pollution control revenue bonds to finance the system under a lease, sublease arrangement. The sale of an additional $39 million of the bonds was sold December 1, 1975, with proceeds to be applied to the remainder of the No. 1 unit system and the AQC system for No. 2 unit.

La Cygne production costs for 1975 for energy averaged 8.82 mills per kilowatt hour. The limestone and operating and maintenance expense of the AQC system totaled 1.26 mills, or about 14 per cent of the production costs. Labor accounted for .45 mills, limestone .55 mills, and other costs were .27 mills.

The total annual cost of producing power including fixed costs from La Cygne No. 1 unit in 1978 is estimated at 15.24 mills per kilowatt hour. Interestingly, the comparative cost for the No. 2 unit, which will burn Wyoming coal supplied by Amax Coal Company from mines in the Gillette area, will be 21.96 mills per kilowatt hour. As these projections indicate, total cost of producing power from No. 1 unit, even with the complicated air quality control system, will still be 44 per cent less expensive than for No. 2 unit. Of the total cost, the air quality component for La Cygne No. 1 unit is 3.03 mills per kilowatt hour, and for No. 2 unit is 1.77 mills per kilowatt hour. Unit No. 2 will be equipped with an electrostatic precipitator.

ADDITIONAL MODIFICATIONS

Planned additional modifications include the installation of an eighth module to be in service the summer of 1977. This is estimated to cost $5.2 million and will improve the cruising capability of the unit from 700 megawatts to 800 megawatts. The original AQC system design allowed space for this modification. Changes in the induced draft fans are also being made to reduce or eliminate a fan paralleling problem. Many other experiments and minor changes are continuing to improve performance and to reduce operating and maintenance expenses.

LA CYGNE STATION
COAL AND ASH ANALYSIS

COAL

Proximate

Volatile	28.63
Fixed Carbon	37.94
Ash	24.36
Moisture	9.07
	100.00

BTU/lb. 9421

Grindability 59.59

Ultimate

Moisture	8.60
Carbon	51.93
Hydrogen	3.43
Nitrogen	0.94
Chlorine	0.027
Sulfur	5.39
Ash	24.36
Oxygen	5.33
	100.007

ASH

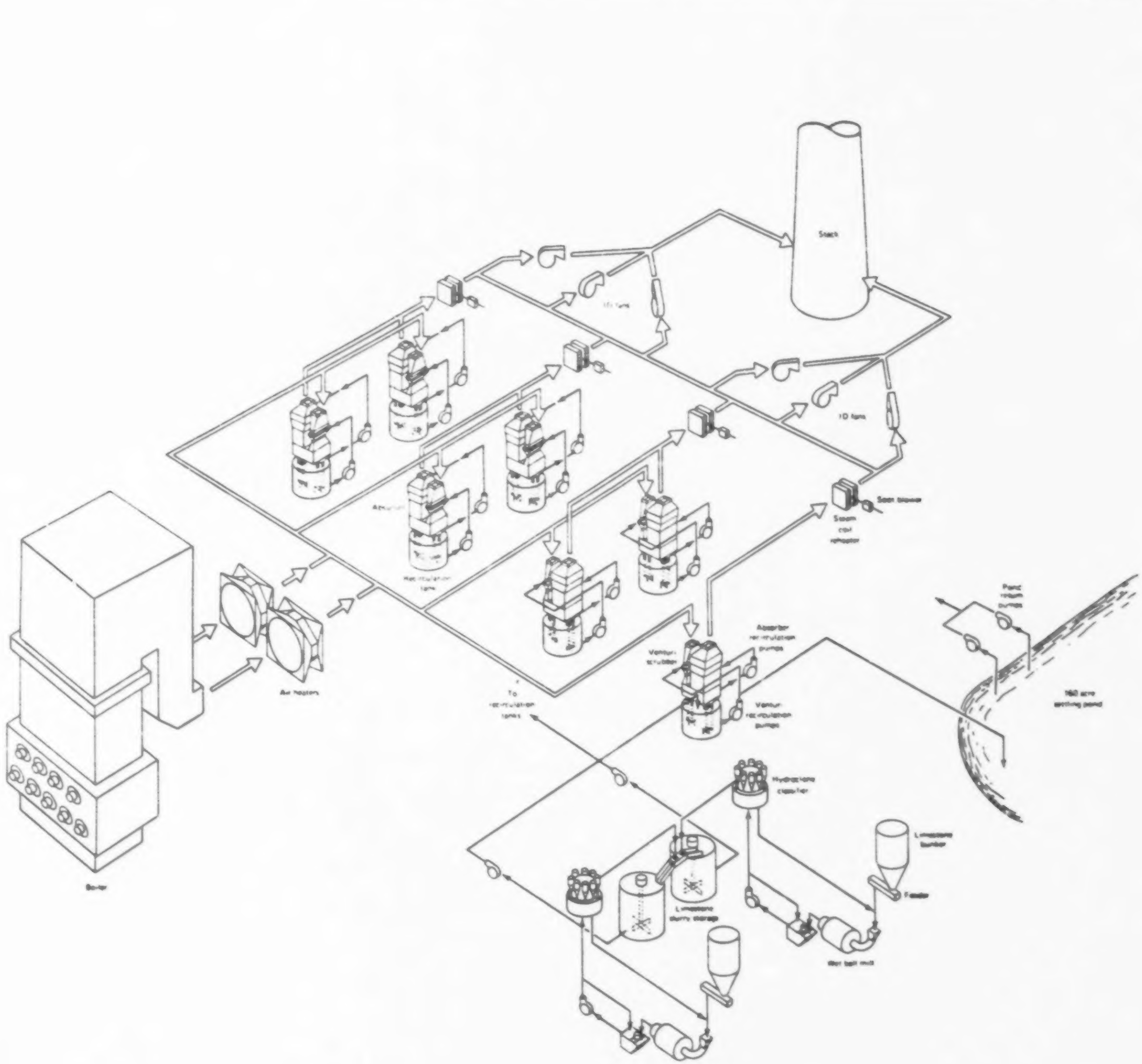
Analysis

Phosphorous Pentoxide	0.15
Silica	46.05
Ferric Oxide	19.23
Alumina	14.07
Lime	6.86
Magnesia	1.02
Sulfur Trioxide	7.85
Potassium Oxide	2.48
Sodium Oxide	0.60
Titania	1.02
Other	0.67
	100.00

Fusion Temperature

Reducing I. D.	1957
Soft (H=W)	2045
Soft (H=W/2)	2169
Fluid	2321
Oxidizing I. D.	2156
Soft (H=W)	2338
Soft (H=W/2)	2415
Fluid	2520

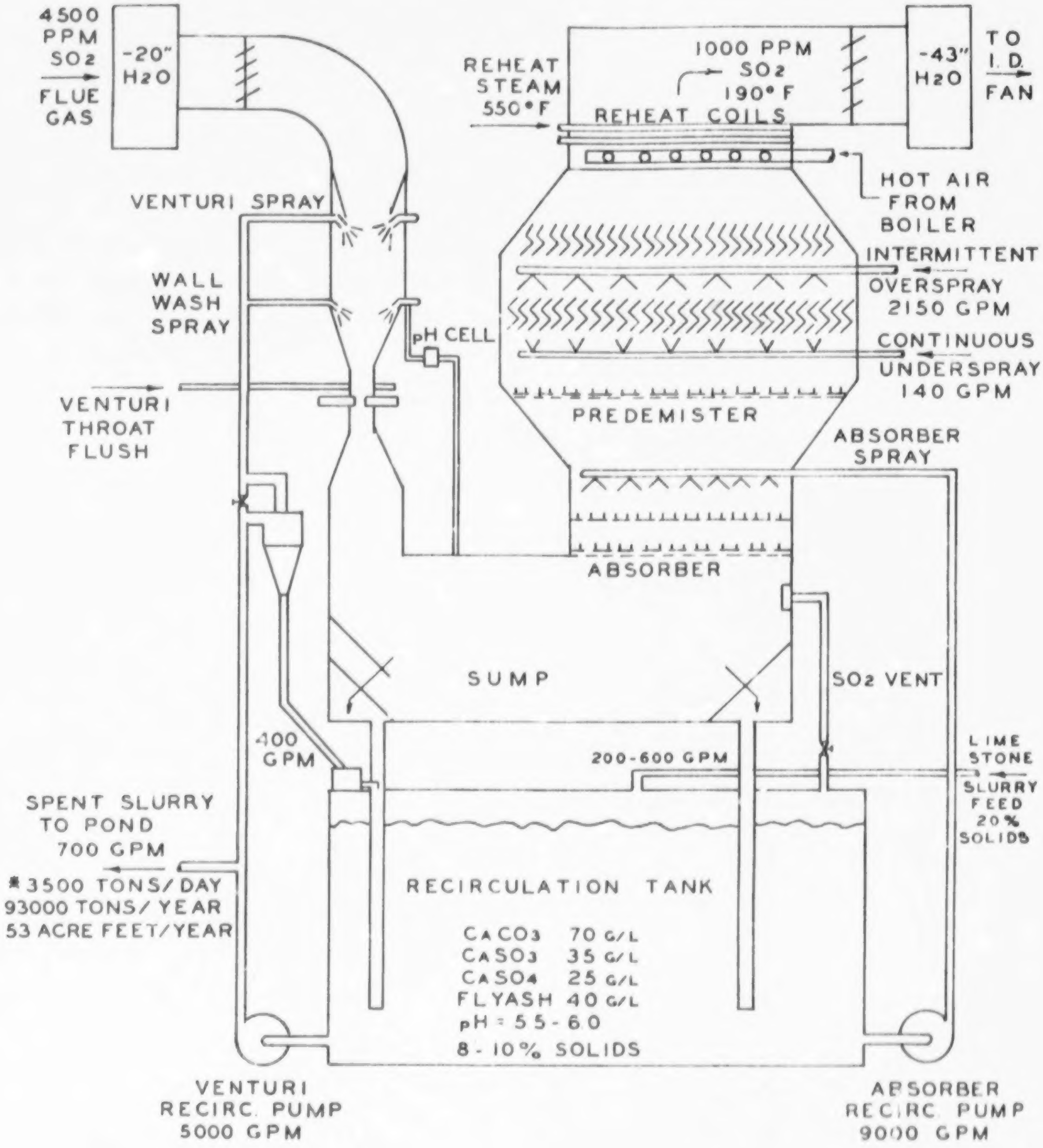
Exhibit A


Stack
ID fans
ID fans
Soot blower
Steam coil reheater
Absorber
Recirculation tank
Air heaters
Boiler
Absorber recirculation pumps
Venturi scrubber
Venturi recirculation pumps
To recirculation tanks
Pond return pumps
160 acre settling pond
Hydroclone classifier
Limestone bunker
Feeder
Wet ball mill
Limestone slurry storage

La Cygne limestone wet scrubbing system

EXHIBIT B

FIGURE 1 - LACYGNE FGD MODULE


4500 PPM SO2 FLUE GAS
-20" H2O
1000 PPM SO2 190° F
REHEAT STEAM 550° F
REHEAT COILS
-43" H2O
TO I.D. FAN
HOT AIR FROM BOILER
VENTURI SPRAY
WALL WASH SPRAY
pH CELL
INTERMITTENT OVERSPRAY 2150 GPM
CONTINUOUS UNDERSPRAY 140 GPM
VENTURI THROAT FLUSH
PREDEMISTER
ABSORBER SPRAY
ABSORBER
SUMP
SO2 VENT
400 GPM
200-600 GPM
LIME STONE SLURRY FEED 20% SOLIDS
SPENT SLURRY TO POND 700 GPM
*3500 TONS/DAY
93000 TONS/YEAR
53 ACRE FEET/YEAR
RECIRCULATION TANK
CaCO3 70 G/L
CaSO3 35 G/L
CaSO4 25 G/L
FLYASH 40 G/L
pH = 5.5-6.0
8-10% SOLIDS
VENTURI RECIRC. PUMP 5000 GPM
ABSORBER RECIRC. PUMP 9000 GPM

*TOTAL FOR ALL MODULES

EXHIBIT C

LA CYGNE SCRUBBER WATER ANALYSIS

CATIONS	COOLING LAKE	SETTLING POND
CALCIUM HARDNESS SOL. (Ca)	123.2	696.0
MAGNESIUM HARDNESS SOL. (Mg)	11.0	48.0
SODIUM (Na)	12.0	22.0
POTASSIUM (K)	4.5	23.0
ANIONS		
BICARBONATE ALK (AS HCO_3)	109.8	36.6
CHLORIDE (Cl)	24.6	177.8
SULFATE (SO_4)	247.9	1627.3
SULFITE (SO_3)	* ND	* ND
SILICA (SiO_2)	2.4	20.6
OTHERS		
pH (pH UNITS)	7.87	7.0
CONDUCTIVITY IN MICROMHOS	649.0	4380.0

*ND - Not Detected

Exhibit D

MODULE AVAILABILITY SUMMARY

LA CYGNE 1975

MONTH	A	B	C	D	E	F	G	TOTAL AVAILABILITY*	MWH	BOILER HOURS	GENERATION CAPACITY FACTORS
January		Turbine Generator Repair									
February		Turbine Generator Repair							7,886		
March	82.4	96.03	89.5	76.6	92.96	91.5	96	89.33	244,873	694	38.8
April		Generator Repair 25 Days							23,014		
May	94.6	85.1	94.2	89.5	89.8	89.3	83.4	89.4	332,526	683	52.7
June	87.8	85.4	83.9	84.9	84.1	86.1	88.6	85.8	324,952	667	53.2
July	78.4	89.7	89.6	83.7	85.4	87.4	85.2	85.6	297,870	590	47.2
August	74.64	88.07	87.29	78.01	92.44	85.00	83.06	84.07	294,402	630	46.7
September	78.43	83.62	84.38	84.67	78.72	77.71	74.24	80.25	239,954	610	39.3
October	66.16	77.26	46.27	73.62	71.91	73.07	64.69	67.57	74,660	231	24.3
		Generator Repair 15 Days									
November	92.87	90.79	80.18	93.18	96.09	89.39	93.94	90.83	165,058	346	50.5
		Generator Repair 17 Days									
December	90.72	87.39	80.87	85.20	86.89	88.56	83.67	86.19	278,597	597	46.8

*$\dfrac{\text{Working Hours + Reserve}}{\text{Hours in Month}}$

Exhibit E

Four Hour Load Test Results

(Test for Maximum Credited Megawatts)

Date: March 24, 1975
Time: 6:10 pm to 10:25 pm
Power: Min. - 800 MW (e) Gross - Peak* - 830 MW (e)
Outside Temp: 26°F

Data

	A	B	C	D	E	F	G
Gas Flows (cfm)	380,000	360,000	400,000	410,000+	400,000	390,000	340,000
Throat Pos.	22"	22"	22"	22"	22"	22"	22"
Reheat Temp. (°F)	170	180	180	150	150	170	185
Absorber Pump	ON	ON	ON	ON	ON	ON	ON
Venturi Pump Flow	4,000	4,000	5,000	4,000	5,000	5,000	5,000
Venturi ΔP	8.0	7.5	-	11.0	10.0	11.0	9.0
Reheater ΔP (X2)	4.0	3.5	3.0	3.5	1.0	4.5	12.5
Abs-Dem. ΔP	-	5.5	-	8.5	8.5	9.0	9.0
Hot Air Damper Pos. (% Open)	46	23	44	-	48	32	64
Reheat Outlet Damp. Pos. (% Open)	51	37	98	84	100	96	100
Scrubber Outlet Press. (-"H_2O)	38.3(common)			34.0			
I.D. Fan Amps	540	520	620	600	380	380	
I.D. Fan Inlet Damper Pos. (% Open)	100	98	98	100	100	23	
F.D. Fans Amps	540	500	490				
Lab pH**	5.95	5.89	5.80	5.68	5.92	5.79	5.73
Conductivity**	2100	2200	2000	2250	2000	2100	2200
Sulfite (g/l)	41.1	36.2	45.9	35.4	45.1	29.4	54.7
Carbonate (g/l)	94.4	84.1	69.4	84.4	91.9	109.0	76.6
SO_2 Efficiency	75.1	79.1	72.2	72.0	?	82.9	75.7
Inlet (ppm)***	5700	5138	5516	4995	5700	5017	5120
Outlet (ppm)	1419	1075	1533	1220	?	857	1243

Exhibit F

COSTS

LA CYGNE STATION

Scrubber Operating Expense 1975

OPERATING LABOR	$ 601,029	0.265 Mils/KWH
OPERATING MATERIALS	195,926	.086
MAINTENANCE LABOR	416,206	0.184
MAINTENANCE MATERIALS	386,397	0.171
LIMESTONE	1,256,048	0.554
TOTAL	2,855,606	1.260 Mils/KWH

CAPITAL COSTS

TOTAL UNIT	$200,000,000	$244/KW
SCRUBBER PORTION	43,000,000	$ 52/KW

COAL: $7.10 TON - 37.7¢ MKB
USAGE 2 to 2.4 TONS PER YEAR (10^6)

LIMESTONE: $2.717 TON (3/4" x 0") APPROX 700,000 TONS/YEAR
29% OF TOTAL COAL USED.

Exhibit G

Module Evaluation
Test # 4

LA CYGNE STATION
STACK SAMPLING TEST

DATE: 5/15/75
TIME: 11:30 P.M. to 8:00 P.M.
LOAD RANGE: 700 to 720 MW Continuous
AMBIENT TEMP: 72° F
AVERAGE SO_2 REMOVAL: 80%
PARTICULATE REMOVALS: 98.2%
BAROMETRIC P.: 29.25
REL. HUM. ________

(Data taken by independent test group)

Inlet 9.9# per 10^6 BTU
Outlet .18# per 10^6 BTU

DATA LOGGED AT 705 MW

	A	B	C	D	E	F	G
Gas Flow Indicated	320K	280 K	300 K	340 K	320 K	300 K	300 K
Throat Position	22	22	22	22	22	22	22
Reheat Temperature	170	190	170	148	180	160	190
Absorber Pump	On	On	On	On	On	On	On
Venturi Slurry Flow	4,000	4,000	4,000	4,000	4,000	4,000	4,000
Venturi Δ P	7	6	7	7	8	7	8
Reheater Δ P	4	3	3	8	6	11	8
Absorber Dem. Δ P	-	4	-	8	8	8	8
Hot Air Damper Pos. % Open	46	20	50	-	50	94	100
Reheat Outlet Damper Pos. % Open	40	35	55	75	82	100	100
Scrubber Outlet Press. (-"H_2O)	38	-	-	28	-	-	-
I.D. Fan Amps (Control Room)	460	480	460	400	400	440	
I.D. Fan Amps (Breaker)							
I.D. Fan Inlet Dampers Pos. % Open	60	60	60	60	60	32	
F.D. Fan Amps (Control Room)							
F.D. Fan Amps (Breaker)							
Lab pH/C.R.	5.95	6.02	5.93	5.80	5.98	5.90	5.90
Sulfite (g/l)	53.9	37.0	37.4	58.4	61.6	54.7	64.8
Carbonate (g/l)	103.0	85.3	108	75.3	109	93.1	95.9
SO_2 Efficiency %	76	81	81	82	77	83	81
Inlet (ppm)	4,506	4,297	4,663	4,273	4,982	4,156	4,834
Outlet (ppm)	1,068	834	892	776	1,121	704	917

BOILER DATA:

F. D. Fan Discharge ("H_2O) 44
Air Flow % 68
Fuel Flow % 72
Feedwater Flow #/Hr. 5,200,000
Excess O_2 % 2.5
Windbox-Furn. Diff. Press. 31
Furn. Pressure -2.0
Sec. Super Gas Pressure -3.4
Pend. Reheat Gas Pressure -3.2
Primary Super Gas Pressure -6.0

Horz. Reheat Gas Pressure -7.3
Econ. Outlet Gas Pressure -10.0
Feedwater Pressure 4,100
Throttle Pressure 3,500
Throttle Temperature 970
Hot Reheat Temperature 1,000°F
Air to Air Heater A 155.5
Air to Air Heater B 156.9
Air from Air Heater A 561.0
Air from Air Heater B 598.0
Gas to Air Heater A 655.5
Gas to Air Heater B 668.7
Gas from Air Heater A 308.7
Gas from Air Heater B 303.9

Exhibit H

LA CYGNE STATION

AMBIENT MONITORING SYSTEM

	STATION 1	STATION 2	STATION 3
MILES FROM PLANT	2	10	12.5
PRIOR TO START UP			
CONCENTRATION SO_2 - ppm	.009	.008	.003
NO_2 - ppm	----	.009	-----
RECENT LEVELS			
CONCENTRATION SO_2 - ppm	.019	.028	.029
NO_2 - ppm		.024	
NATIONAL STANDARDS			
CONCENTRATION SO_2 - ppm	.030 (80mg/M^3)	.030	.030
NO_2 - ppm		.050 (100mg/M3)	

LEAD PEROXIDE CANDLES

21 CANDLES RING THE STATION AT DISTANCES FROM 2 to 6 MILES. 3 MORE CANDLES ARE LOCATED ABOUT 30 MILES NORTH OF THE PLANT, JUST SOUTH OF THE KANSAS CITY METROPOLITAN AREA.

SULFATION GRADUALLY INCREASED FROM .029 TO .031 mg/CM^2 PRIOR TO COMMERCIAL PLANT OPERATIONS AND HAS SINCE INCREASED TO .082 mg/CM^2.

Exhibit I

LA CYGNE AIR QUALITY CONTROL
MANPOWER REQUIREMENTS

OPERATORS PER SHIFT

3 Attendants	13
3 Clean-Up	14
1 Shift Foreman	5
1 Process Attendant (Chemist)	1
	33

MAINTENANCE

Mechanics	8
Apprentice Mechanics	2
Welder	1
Electrician	1
Technician	1
Plant Helpers	2
Foreman	1
	16

ADMINISTRATIVE

Superintendent	1
Engineer	1
	2
TOTAL	51

Exhibit J

RECENT SCRUBBER EXPERIENCE AT THE LAWRENCE ENERGY CENTER
THE KANSAS POWER AND LIGHT COMPANY

D. M. Miller, Manager, Electricity Production

Kansas Power and Light Company
818 Kansas Avenue
Topeka, Kansas 66612

Recent Scrubber Experience at the Lawrence Energy Center
The Kansas Power and Light Company

I would like to share with you our participation in the practical application of the Combustion Engineering Air Pollution Control Systems installed on two steam generating units located at The Kansas Power and Light Company Lawrence Energy Center. These steam generators were both supplied by Combustion Engineering and are capable of firing pulverized coal, fuel oil, or natural gas, each or all together.

My Company is The Kansas Power and Light Company headquartered in Topeka, Kansas. We operate wholly in the State of Kansas. Our peak load this past summer was just short of 1½ million KW. Our principal electric generating stations are located at Tecumseh, Hutchinson and Lawrence, Kansas. Over the past 25 years we have primarily burned natural gas supplemented by fuel oil and/or coal when the gas supply has been interrupted. These interruptions have gone from 8-12% in the 50's and 60's to over 50% in 1975 and we would expect these interruptions co exceed 65% this year. This shift from gas to fuel oil and coal is not a shock and is an occurance we had planned for - it is just here a little earlier than we had programmed.

Even though in 1967, 65% of our steam generators were equipped to burn pulverized coal, we were still classified as a gas fired utility. In 1967 when the decision was made to build the 400 MW addition at our Lawrence Energy Center, it was quite clear that natural gas would not be available to fulfill its fuel requirement. A 20 year contract for Kansas coal with delivery starting in 1967 was negotiated to fulfill the primary fuel requirement for this unit. Natural gas and fuel oil would supplement the coal burning the first years of operation after start up in the spring of 1971.

Also in 1967 when the decision to build this coal burning addition was made, it was assumed that by 1971 there would be some ambient and/or emission regulations in effect for particulate matter and sulfur dioxide. Based on this assumption and the availability of 3-4% sulfur, 12% ash coal, the decision was made to install as original equipment, facilities to treat the stack gas to remove both fly ash and sulfur dioxide. This was accomplished by the installation of a Combustion Engineering Limestone Injection-Wet Scrubber System.

Since the startup of the 400 MW unit was scheduled for 1971, it was further decided to retro-fit a similar system on an existing 125 MW unit at Lawrence so that we could gain some operating experience before starting up the larger unit. This smaller unit went into service in the fall of 1968 and has provided the industry with many innovations which are now being used on newer installations being planned in the United States.

The Combustion Engineering Air Pollution Control System at Lawrence is designed to remove both fly ash and sulfur dioxide simultaneously from the flue gas stream. Both the scrubbing of the fly ash and the chemical reaction to absorb the SO_2 are accomplished in a common vessel.

I would like to describe for you the system as it was initially installed on our 125,000 KW unit. (Figure 1)

The limestone injection equipment is the existing pulverizers where the rock and coal are pulverized together and are blown into the furnace as a mixture. The wet scrubber system contains the spray nozzles, the marble bed, the demisters and reheater coils. The heat source for the reheater is extraction steam from the turbine through the existing feed water heater stream. New I.D. fans were installed to handle the increase in head requirements.

The actual system has two one-half capacity scrubbers each with its own inlet duct and damper, bypass duct and existing stack.

The bypass system was installed so that we could keep the steam generator in service on fuels other than coal so that we could maintain reliability of KW output while learning how to operate the Air Pollution Control System. We have been able to do just that as we have operated many hours with scrubbers out of service. This has allowed us much opportunity to maintain and modify this system while generating KWH which still is our primary job.

The dirty pot overflow water from both scrubbers containing about 2% solids is pumped to the settling pond where the solids are separated. The clarified water is returned through an overflow structure to the clear side of the pond and to the scrubber supply pump intake structure. These pumps supply water to the scrubber system and the cycle repeats. The supply pumps and ponds will be common to the 125 and 400 MW units.

The scrubber on Unit 4 was started up in 1968 with the configuration shown in Figure 2. This was a simple system but presented many operating problems and shortcomings:

Problems encountered were:

1. Build up and plugging of the inlet duct where hot gases entered the scrubber.
2. Erosion of the scrubber walls.
3. Corrosion of the scrubber internals
4. Plugging and scaling of drain lines, tanks and pumps
5. Plugging and scaling of the marble bed
6. Plugging of spray nozzles
7. Plugging of demister due to carryover from bed
8. Plugging of reheater due to carryover through demister
9. Buildup on ID Fan rotors due to moisture getting through reheater under abnormal conditions resulting in fan unbalance.

In addition to the operating problems the SO_2 removal was quite low due to the overburning of the limestone and the dropout of the lime with the ash in the bottom of the scrubber. Satisfactory particulate removal was achieved however.

After the first winter's operation, the scrubber was revised to the configuration shown in Figure 3. The addition of soot blowers in the inlet duct and reheater helped the plugging problems there. The demister was raised to increase the distances between the bed and demister. The pot overflows were directed to the storage pond and a large recycle tank and pump were installed to accumulate and pump the highly alkaline bed underflow back into the bed. New type spray nozzles were installed to minimize plugging. The bottom part of the scrubber tanks were lined with gunite to reduce erosion and corrosion of the walls. We borrowed the coal pulverizers from an adjacent coal fired unit and pulverized the limestone in one of these mills and blew the additive material into the furnace on through the observation doors at the arch level of the boiler. This did provide a reactive material for use in the scrubbers.

Most of the problems were reduced but not eliminated by the first set of revisions. The recirculation system did improve the SO_2 removal. Maximum SO_2 removal was possible under this mode of operation but the resulting sulfur scale formations in the bed drain lines and pumps would soon put the scrubber out of business.

Revisions were made to the scrubbers the following summer in an attempt to further minimize the corrosion, erosion, scaling and plugging problems. Figure 4 shows the configuration of the scrubber in October, 1970. Major revisions made were:

1. Sandblasting and coating the interior of the scrubber vessels with a two coat glass flake lining.
2. Replacement of all internal steel pipe systems with plastic and fiberglass.
3. Replacement of stainless steel demisters with fiberglass demisters.

4. Addition of a ladder vane under the bed to improve gas distribution.

5. Modification of the pot overflow drain piping to allow the drains to return to the recycle tank allowing for a semi-closed loop operation.

6. The original copper fintube reheat coils were removed and replaced with a carbon steel fintube coil. The copper units plugged easily and the fins were flattened by the soot blower jets.

7. We replaced the 14" drain outlets from each scrubber with a 2' x 4' sluice way to alleviate the drain pluggage we had been experiencing.

All the above modifications helped in reducing operating problems except for the demister plugging. Scaling could be controlled by maintaining close ph control but only at the expense of reduced SO_2 removal. Manual washing of the demisters was necessary on an every other night basis to maintain load capability of the unit.

We operated the scrubber in this manner with a fair amount of success through the winter of 1970-71 and into the spring of 1971. We were able to by frequently inspecting the system and rather continually performing small maintenance jobs, keep the scrubber system available to meet the required coal burning during this period.

We were not satisfied with the reliability of the system at this time but planned to start up the 400 MW Lawrence addition under this mode of operation. I will discuss that operation later in this report.

We continued our fight to alleviate the nagging pains of clogging the recycle nozzles or the recycle piping system and more carryover from the bed than the demister could reasonably handle which plagued us with a dirty reheater which did not lend itself to simple repeated cleaning.

Then came the operation of the winter of 1971-72 when we were also burning coal on the 400 MW Lawrence addition.

The only thing we could consistently do was make sulfur bearing scale -- on the scrubber walls -- in the scrubber marble bed and on the demisters of all the scrubbers in both units.

Operation of the scrubber systems remained in this mode until the late winter of 1972. At that time Combustion Engineering completed tests in their laboratory scrubber at Windsor using a high solid slurry crystallization process to control saturation and precipitation to eliminate scale within the scrubber. The system was quite successful in the lab so a crash program was started to modify the Lawrence 4 scrubber for full scale testing of the high solids system.

The temporary revamp took about 30 days to complete and the scrubber was tested for 60 days during May-June, 1972 using the high solids slurry system to control scaling. The results were encouraging; so encouraging in fact that the decision was made to spend the summer of 1972 to rebuild and modify the scrubber systems on both Lawrence 4 and 5 to operate with the high solid-crystallization concept.

The modification included the installation of:

1. Large crystallization tank
2. 4' enlargement of the scrubber bed on #4
3. New rubber lined pipe slurry system
4. New plastic spray nozzles
5. New bed drain system
6. New slurry pumps and strainers
7. New multiple mixers in tank
8. New two bank fiberglass demister with a power wash system

This mode of operation is shown on Figure 5.

The operation of this system on Unit 4 through 1973 was much improved over anything we had been through in the past. We still experienced some of the nagging maintenance and cleaning problems such as:

1. Isolated corrosion
2. Unsatisfactory damper operation
3. Expansion joint failure

4. Demister fouling

5. Rapid slurry pump wear

6. Valve failures

but were able by the end of 1973 to maintain the scrubber system operable with 2 - 8 hour shifts of manual cleaning per scrubber per week.

This cleaning requirement lessened a little into 1974 when for economical reasons in the cost of coal, we completed negotiations for a supply of low sulfur coal from SE Wyoming for these units. We started receiving this coal: 10,000 BTU/#, .5% sulfur and 11% ash in the fall of 1974 and had phased out all the SE Kansas coal by late spring 1975. The bypass duct system in the Lawrence 4 unit was removed at this time due to complete deterioration.

The operation of the scrubber system on the Wyoming coal has proved to be somewhat better and more economical due to a lesser amount of sulfur removal required. The scrubber system is still operating in the high solids mode as an SO_2 and particulate removal system. Our normal manual cleaning requirements have been reduced to 2-4 hour shifts per scrubber per week.

In 1974 this unit was available for operation 343 days. 50% of the fuel consumed was coal, 2% fuel oil and 48% natural gas.

In 1975 this unit was available for operation 333 days. 64% of the fuel consumed was coal, 3% fuel oil and 33% natural gas.

I would estimated this unit to be available for operation 330 days in 1976 with the following fuel split; 80% coal, 8% fuel oil and 12% gas.

Remember the scrubber system must be in operation to burn coal.

You may have heard that we are replacing this scrubber system with a rod section followed by a spray tower and wonder why. You have a right to that question.

We are presently replacing this successful scrubber system because since 1968 in modifying and revising the scrubber modules and operating many hours at corrosion levels that were bad, we actually consumed the physical scrubber plant. Now that we have achieved - at the Lawrence Energy Center and under the conditions of operation there - a mode of operation that we can maintain, we have had so much deterioration of vessel and equipment that we must build a new system to have one.

We are extremely interested in the new system since parts of this system will be installed at our new Jeffrey Energy Center which will be discussed later.

So much for the Lawrence #4 Unit, lets now review the operations of the Lawrence #5, 400,000 KW unit - installed in 1971.

We followed the programs as discussed in Lawrence 4 with the scrubber system on this unit. We started with recycle to the bed - we made scale, we modified and installed the high solids mode of operation, we added 2 scrubber modules and we operated burning coal when it was required.

We experienced the problems of Lawrence 4 on this unit with one additional problem, a more severe distribution of flue gas to and through the marble beds on #5 than on #4. This has complicated our lives on #5 unit and has been a large part of the reason our operating successes on #5 have come slower and not reached as high a plane as #4.

However we achieved the following performance with Unit #5.

In 1974 - the unit was available for operations 338 days - consuming 27% coal, 6% fuel oil and 66% gas.

In 1975 - the unit was available for operations 352 days - consuming 42% coal, 13% fuel oil and 45% gas.

I expect this unit to be available for service 330 days in 1976 - consuming 60% coal, 25% fuel oil and 15% gas.

We are studying the future of the scrubber system on Unit 5 and my best guess today is that it will follow in the footsteps of Unit 4 and will be converted to the rod and spray tower system.

These experiences at our Lawrence Energy Center are the basis for our conclusions concerning the stack gas clean-up system we intend to install at the Jeffrey Energy Center. (Figure 6)

The Jeffrey Energy Center is composed of 4 - 680,000 KW coal fired units planned to be on line 1 unit early 1978, 1 in 1980, 1 in 1982 and 1 in 1984 All of these units will be fired by coal now under contract from the AMAX Coal Company from the Powder River Basin in Wyoming. The coal specifications for this plant are designed to be 8000 BTU/#, .32% sulfur and 5.8% ash.

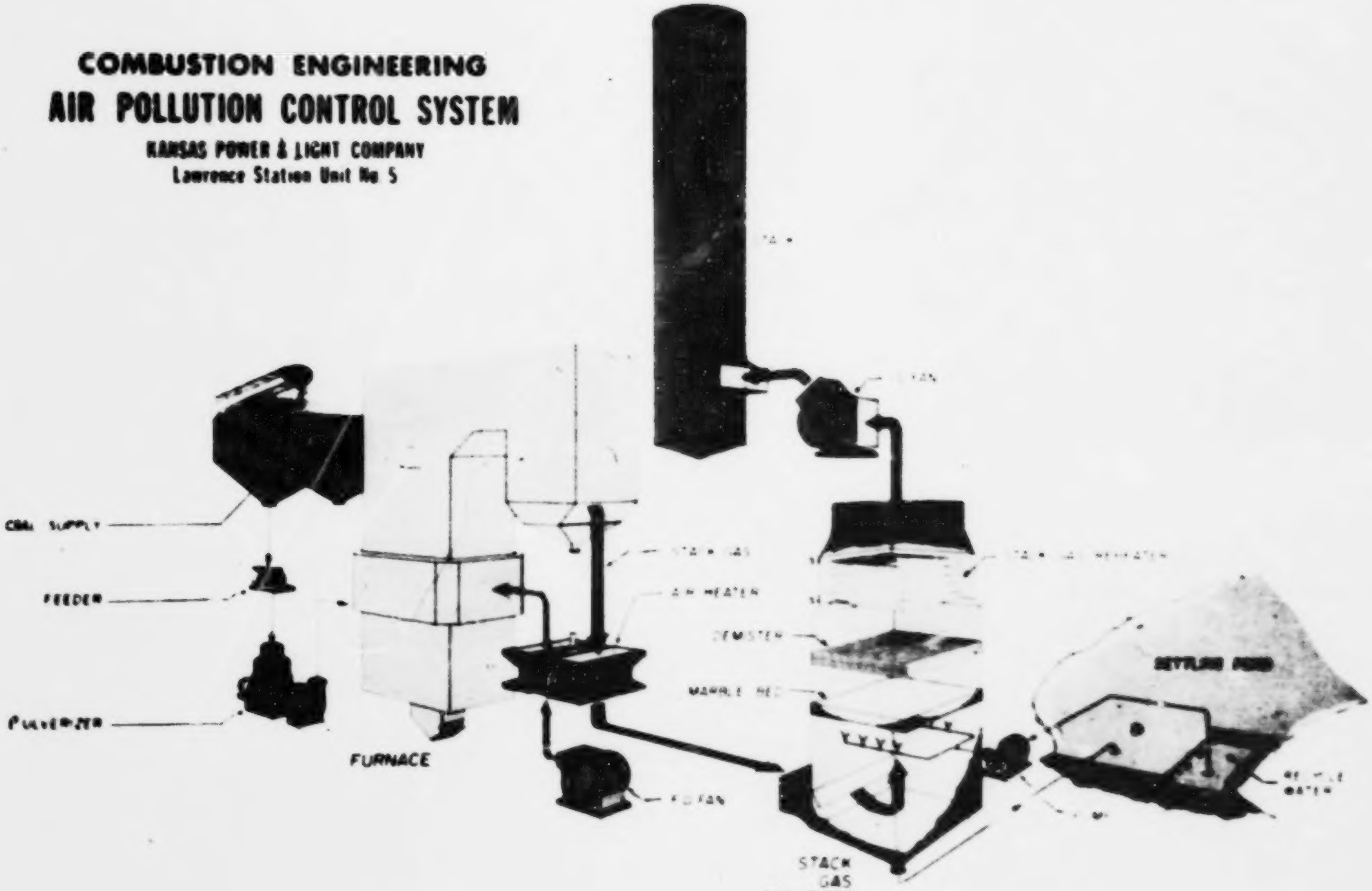
We have purchased from Combustion Engineering a stack gas clean-up system that will remove in excess of 99% of the particulate matter, and 60% of the SO_2 in the flue gas. An overfire air system at the tangential fired pulverized burners will control the Nox emission.

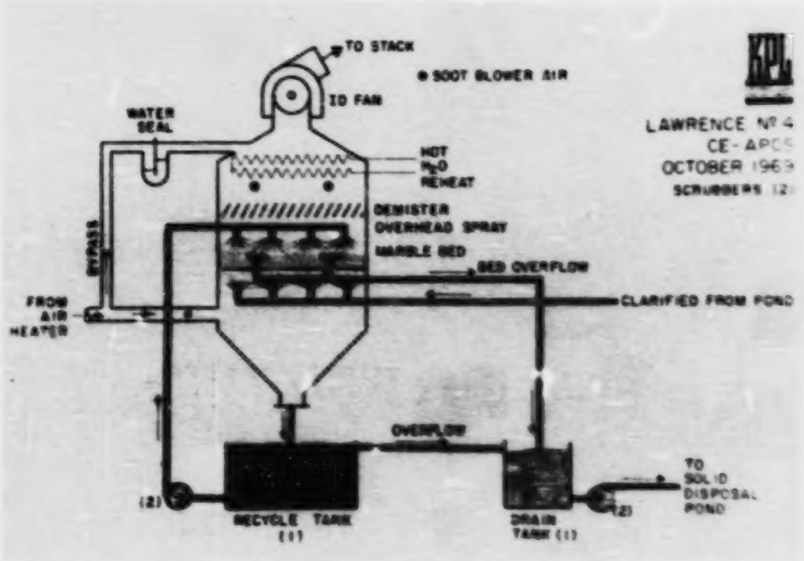
The stack gas clean-up system (Figure 7) is composed of an electrostatic precipitator following the air heater then the induced draft fans and then through spray towers. The effluent gas from the spray towers will be mixed with some hot gas from the ID fan discharge for reheating the flue gas as it flows to and out the 600 ft. chimney.

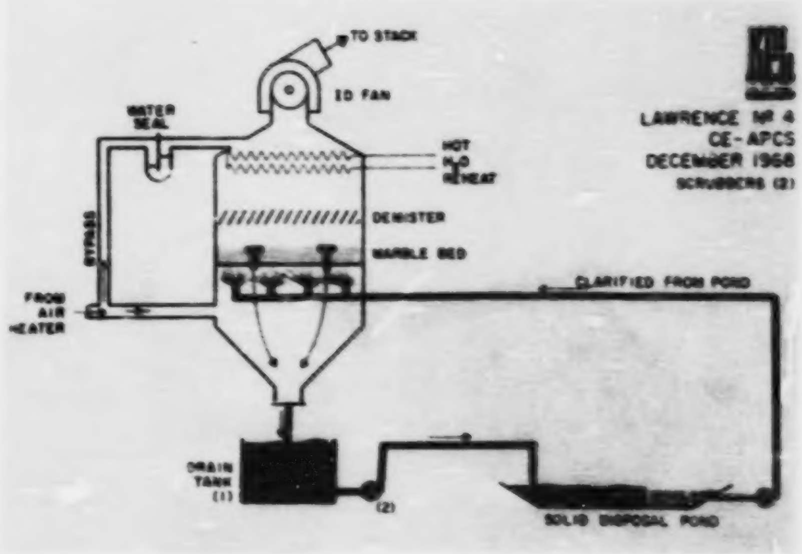
We are of the opinion that we will be able to operate this system under the conditions we will have at the Jeffrey Energy Center in such a manner to be good neighbors to those who live around us in Pottawatomie County, Kansas and to satisfy the existing requirements of The Department of Health and Environment of the State Of Kansas and the Environmental Protection Agency.

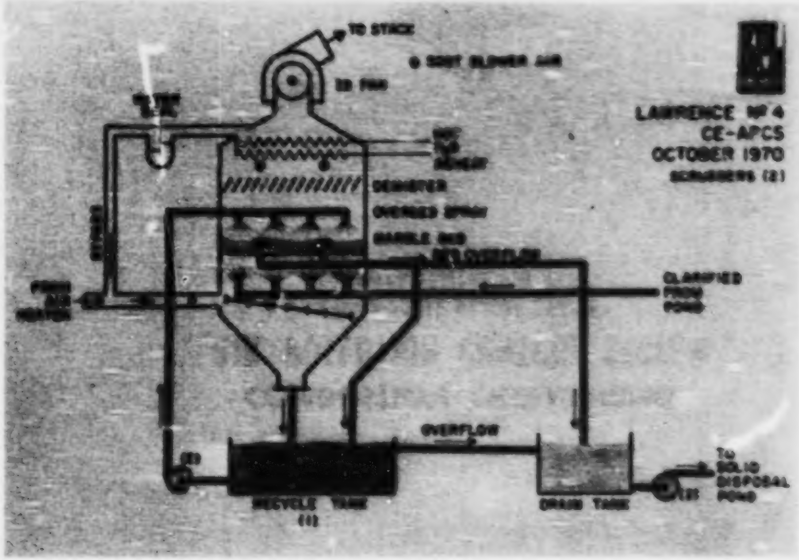
COMBUSTION ENGINEERING
AIR POLLUTION CONTROL SYSTEM

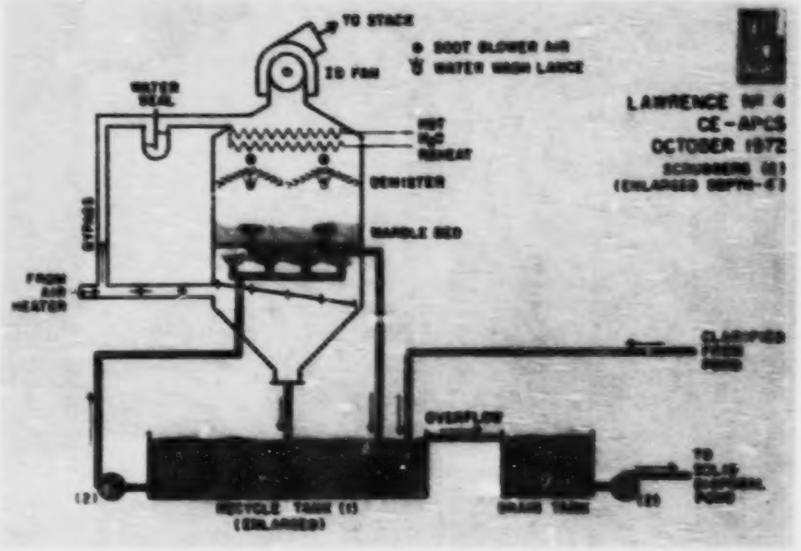
KANSAS POWER & LIGHT COMPANY
Lawrence Station Unit No 5

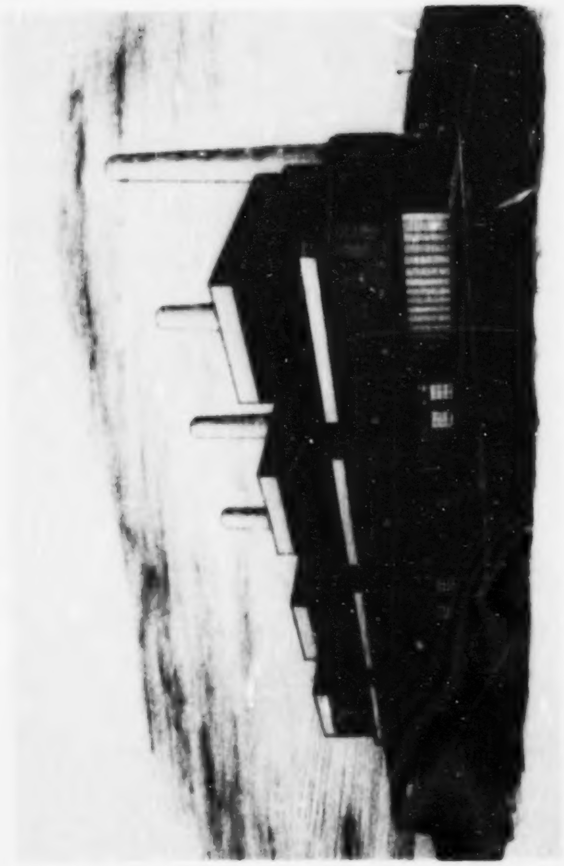

STACK
I.D. FAN
COAL SUPPLY
STACK GAS
STACK GAS REHEATER
FEEDER
AIR HEATER
DEMISTER
SETTLING POND
MARBLE BED
PULVERIZER
FURNACE
F.D. FAN
RECYCLE WATER
PUMP
STACK GAS SCRUBBER

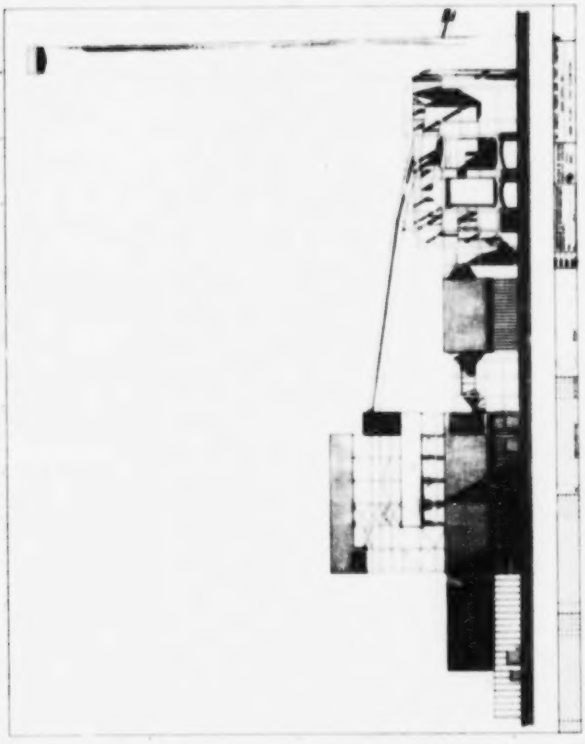

TO STACK
● SOOT BLOWER AIR
ID FAN
WATER SEAL
LAWRENCE Nº 4
CE-APCS
OCTOBER 1969
SCRUBBERS (2)
HOT
H_2O
REHEAT
DEMISTER
OVERHEAD SPRAY
MARBLE BED
BED OVERFLOW
BYPASS
FROM AIR HEATER
CLARIFIED FROM POND
OVERFLOW
(2)
RECYCLE TANK (1)
DRAIN TANK (1)
(2)
TO SOLID DISPOSAL POND


TO STACK
ID FAN
WATER SEAL
LAWRENCE Nº 4
CE-APCS
DECEMBER 1968
SCRUBBERS (2)
HOT
H_2O
REHEAT
DEMISTER
MARBLE BED
BYPASS
FROM AIR HEATER
CLARIFIED FROM POND
DRAIN TANK (1)
(2)
SOLID DISPOSAL POND


TO STACK
● SOOT BLOWER AIR
ID FAN
WATER SEAL
LAWRENCE Nº 4
CE-APCS
OCTOBER 1970
SCRUBBERS (2)
HOT
H_2O
REHEAT
DEMISTER
OVERHEAD SPRAY
MARBLE BED
BED OVERFLOW
BYPASS
FROM AIR HEATER
CLARIFIED FROM POND
OVERFLOW
(2)
RECYCLE TANK (1)
DRAIN TANK
(2)
TO SOLID DISPOSAL POND


TO STACK
● SOOT BLOWER AIR
WATER WASH LANCE
ID FAN
WATER SEAL
LAWRENCE Nº 4
CE-APCS
OCTOBER 1972
SCRUBBERS (2)
(ENLARGED DEPTH-6')
HOT
H_2O
REHEAT
DEMISTER
MARBLE BED
BYPASS
FROM AIR HEATER
CLARIFIED FROM POND
OVERFLOW
(2)
RECYCLE TANK (1)
(ENLARGED)
DRAIN TANK
(2)
TO SOLID DISPOSAL POND





INTRODUCTION TO DOUBLE ALKALI FLUE GAS DESULFURIZATION TECHNOLOGY

Norman Kaplan

Industrial Environmental Research Laboratory
Office of Research and Development
Environmental Protection Agency
Research Triangle Park, North Carolina 27711

ABSTRACT

The chemistry and significant process and design factors applicable to sodium/calcium double alkali systems are presented. Technical terminology associated with these systems is defined.

Estimates of capital and operating costs are presented in addition to reasonable targets for reagent consumption in double alkali systems.

The EPA development program is summarized and a status of technology is presented.

Preceding page blank

NOTES

1. Company Names and Products.

   The mention of company names or products is not to be considered an endorsement or recommendation for use by the U.S. Environmental Protection Agency.

2. Consistency of Information.

   The information presented was obtained from a variety of sources (sometimes by telephone conversation) including system vendors, users, EPA trip reports and other technical reports. As such, consistency of information on a particular system and consistency of information between the several systems discussed may be lacking. The information presented is basically that which was voluntarily submitted by developers and users with some interpretation by the author. The order of presentation of information or the amount of information presented for any one system should not be construed to favor or disfavor any particular system.

3. Units of Measure.

   EPA policy is to express all measurements in Agency documents in metric units. When implementing this practice will result in undue cost or difficulty in clarity, IERL/RTP is providing conversion factors for the particular non-metric units used in the document. Generally, this paper uses British units of measure.

   For conversion to the Metric system, use the following equivalents:

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 meter
1 ft^2	0.0929 $meters^2$
1 ft^3	0.0283 $meters^3$
1 grain	0.0648 gram
1 in.	2.54 centimeters
1 $in.^2$	6.452 $centimeters^2$
1 $in.^3$	16.39 $centimeters^3$
1 lb (avoir.)	0.4536 kilogram
1 ton (long)	1.0160 metric tons
1 ton (short)	0.9072 metric tons
1 gal.	3.7853 liters

INTRODUCTION AND BACKGROUND

At the last EPA Symposium on Flue Gas Desulfurization (FGD), held in Atlanta in November 1974, plans were announced for an EPA co-funded full-scale utility boiler double alkali (D/A) demonstration program in which the system vendor and utility would be selected on a competitive basis. At that time there were two industrial boiler applications and an industrial kiln control application of the technology in this country. Japan was a bit more advanced in the application of this technology in that there was one 150 Mw application of the technology in a full-scale utility system. As a result of the increased testing of D/A systems in this country in various pilot plants, at a 20 Mw utility prototype system at Gulf Power, at various industrial boiler systems and other industrial applications (sulfuric acid plants, kilns,etc.) and the interest generated by EPA's request for proposal (RFP) for a full-scale utility demonstration of D/A technology, this technology now appears to be competitive with lime/limestone wet scrubbing for some utility applications.

In response to the RFP, EPA received three viable proposals for installation of double alkali systems at full-scale coal-fired utility systems, ranging in size from 192 to 575 Mw.

In Japan there are now three operational full-scale utility applications ranging in size from 150 to 450 Mw.[18] Additionally, Kureha has developed another variation of the D/A process which, reportedly, is a significant improvement over their limestone D/A system that they now have operating as a full-scale utility system. This variation is called the "Sodium Acetate-Gypsum" process and uses sodium acetate as the SO_2 absorbent, limestone as the regenerant, and produces a gypsum by-product. This process is currently being pilot tested at a 5000 Nm^3/hr (1 Mw) pilot plant using a perforated tray tower operating at a 280 mm (11 inches) H_2O pressure drop, L/G of 7-8 gal./1000 ft^3 gas with an SO_2 inlet concentration of 1400 ppm and an exit concentration of less than 10 ppm. This process will be discussed in greater detail in a later paper in this session.

"Double alkali" or "dual alkali" processes as they have come to be known, like their precursors the lime or limestone wet scrubbing processes, are aqueous alkali scrubbing processes used for FGD. If a "black box" view of these processes is employed, they are, with minor exception, like the lime/limestone scrubbing systems:

lime or limestone is consumed, and a calcium sulfite/sulfate and fly ash wet solid waste product is produced. A more detailed examination inside the "black box," however, would show that the overall process has been split into a number of intermediate steps designed to improve upon the lime/limestone processes by increasing reliability of operation, utilization of lime and/or limestone, and sulfur oxide removal efficiency and, under certain circumstances, producing a solid waste with better handling characteristics. Whereas in a lime/limestone process the absorption of the SO_x from the flue gas and the production of the waste product occur to some extent simultaneously in a single reaction system, in the double alkali processes, these two steps are separated through the use of an intermediate soluble alkali; absorption and production of waste product can then occur in separate system components.

Separating the absorption and waste production functions accomplishes two very important objectives First, it permits scrubbing the flue gas with a soluble alkali thus limiting the SO_x absorption reaction only to gas/liquid chemical equilibrium and to the rate of transfer of SO_x from the flue gas to the scrubbing solution. In lime/limestone processes the rate of dissolution of lime or limestone is a third important limiting factor. Thus SO_x absorption efficiency in a double alkali system is potentially higher than in a lime/limestone system with the same physical dimensions and liquid/gas flow rates. Additionally, soluble calcium is minimized and calcium slurry is kept out of the scrubbing apparatus thus preventing solids scaling and plugging in this critical area. Another benefit of this is better control of the lime/limestone reaction with acidic sulfur containing compounds in separate equipment specifically designed for this reaction thus potentially increasing lime/limestone utilization.

Although a number of other processes can technically be considered double alkali processes, this paper is limited to consideration of the sodium/calcium based double alkali processes. In these processes, a soluble sodium based alkali ($NaOH$, Na_2SO_3, Na_2CO_3, $NaHCO_3$) is used to absorb SO_x from the flue gas in the scrubber, and then a calcium based alkali ($Ca(OH)_2$, CaO, $CaCO_3$) is reacted with the SO_x-rich scrubber effluent liquid to precipitate the insoluble $CaSO_3 \cdot 1/2\ H_2O$, $CaSO_4 \cdot 2H_2O$ and mixed crystal and regenerate sodium based soluble alkali for recycle to the scrubber system. Double alkali systems with an ammonium/calcium base have been tested: while they might have advantages in the reaction with calcium compounds, their main disadvantage is the potential for pollution by a visible ammonium salt plume from the scrubbing apparatus caused by the highly volatile ammonium compounds. Another variation that might be considered double alkali is Monsanto's "Calsox" process, which uses an aqueous organic base as the absorbent solution in combination with lime as the calcium-supplying alkali to produce the throwaway product.

DEFINITION AND DISCUSSION OF TERMS

As with any specialized technology, a discussion of flue gas desulfurization in general, and double alkali technology more specifically, involves the use of special terminology which has evolved with the technology. While terms are understandable to those dealing with the subject on a daily basis, they can be somewhat ambiguous to others. To clarify some of these ambiguities, and to define terms used here and by others describing double alkali technology development, a number of terms and concepts are defined and discussed in a general sense.

Absorption/Regeneration Chemistry

The main chemical reactions that take place in double alkali systems can be divided functionally into the absorption and regeneration reactions. A number of secondary reactions which have very important effects on the overall functioning of the system also take place. These include oxidation, softening and sulfate removal reactions which are discussed under the appropriate headings.

The regeneration reactions and in some cases the absorption reactions will be dependent upon which calcium supplying regenerant is used--lime or limestone. With lime the system can be operated over a wider pH range than with limestone. This wider pH range allows lime systems to operate over the complete range of active alkali hydroxide/sulfite/bisulfite whereas limestone systems can only operate in the sulfite/bisulfite range.

The main overall absorption reactions are described by the following equations:

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O \qquad (1)$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \qquad (2)$$

The main overall regeneration reactions are described by the following equations for lime and limestone respectively:

Lime

$$Ca(OH)_2 + 2NaHSO_3 \rightarrow Na_2SO_3 + CaSO_3 \cdot 1/2\ H_2O\downarrow + 3/2\ H_2O \qquad (3)$$

$$Ca(OH)_2 + Na_2SO_3 + 1/2\ H_2O \rightarrow 2NaOH + CaSO_3 \cdot 1/2\ H_2O\downarrow \qquad (4)$$

$$Ca(OH)_2 + Na_2SO_4 + 2H_2O \rightleftarrows CaSO_4 \cdot 2H_2O\downarrow + 2NaOH \qquad (5)$$

Limestone

$$CaCO_3 + 2NaHSO_3 + 1/2\ H_2O \rightarrow Na_2SO_3 + CaSO_3\cdot 1/2\ H_2O\downarrow + CO_2 + H_2O \quad (6)$$

D/A systems are frequently and erroneously referred to as sodium ion scrubbing systems. It should be stated that, from a "pure" chemistry viewpoint, the reactions presented in equations (1-11) and (15-18) do not involve the sodium ion (Na+); however, the presentation is made using compounds of sodium because sodium systems are prevalent in D/A applications and because this allows showing the reactions using electrically neutral reactants and products rather than charged ions. For example, the absorption reactions involve reaction of SO_2 with an aqueous base such as OH^-, $SO_3^=$, HCO_3^- or $CO_3^=$ rather than with Na^+ which does not take part in the reaction, but is only present to maintain electrical neutrality. Thus equations (1) and (2) for example could have been written as equations (1a) and (2a) respectively:

$$2\ OH^- + SO_2 \rightarrow SO_3^= + H_2O \quad (1a)$$

$$SO_3^= + SO_2 + H_2O \rightarrow 2HSO_3^- \quad (2a)$$

Active Alkali

This term is the sum of the concentrations of NaOH, Na_2CO_3, $NaHCO_3$, Na_2SO_3 and $NaHSO_3$ in the scrubbing solutions. Sodium bisulfite is included in this definition although it is not technically an alkali (i.e., it cannot react with SO_2 in these systems); however, it can be converted to an alkali by reaction with lime or limestone. It should also be noted that the molar capacity of each of these species for absorption of SO_2 is different, and can vary from zero to 2 moles of SO_2 per mole of active alkali. This difference in molar capacity for absorption of SO_2 is illustrated by the following reaction equations:

$$Na_2CO_3 + 2SO_2 + H_2O \rightarrow 2NaHSO_3 + CO_2 \quad (7)$$

(sodium carbonate molar capacity: 2 moles SO_2/mole)

$$NaHCO_3 + SO_2 \rightarrow NaHSO_3 + CO_2\uparrow \quad (8)$$

(sodium bicarbonate molar capacity: 1 mole SO_2/mole)

$$NaOH + SO_2 \rightarrow NaHSO_3 \quad (9)$$

(sodium hydroxide molar capacity: 1 mole SO_2/mole)

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \quad (10)$$

(sodium sulfite molar capacity: 1 mole SO_2/mole)

$$NaHSO_3 + SO_2 \rightarrow \text{No reaction} \quad (11)$$

(sodium bisulfite molar capacity: zero mole SO_2/mole)

Molar capacity is simply the number of moles of SO_2 needed to convert 1 mole of the absorbent alkali completely to sodium bisulfite. Since there is a difference in the molar capacity of different active alkali components to absorb SO_2, active alkali is a descriptive rather than a quantitative term. If the concentration of each of the active alkali components (moles/liter) is known, however, the capacity of the scrubbing liquor to absorb SO_2 (moles of SO_2/liter of solution) can be calculated as the sum of each of the active alkali component concentrations multiplied by their respective molar capacities as follows:

Scrubber liquor SO_2 capacity (moles/liter) =

$$2\,[Na_2CO_3] + [NaOH] + [NaHCO_3] + [Na_2SO_3]$$

TOS

This is an abbreviation for "total oxidizable sulfur." It denotes the concentration of sulfur compounds in solution in which the sulfur is in the +4 oxidation state. Simply, this is the total concentration of sulfite plus bisulfite.

$$\text{TOS (moles/liter)} = [SO_3^{=}] + [HSO_3^{-}]$$

Sulfate is not part of TOS, since the sulfur is in the +6 oxidation state in this specie. Sulfur dioxide dissolving in scrubbing solutions increases the TOS in solution.

Active Sodium

This is the concentration of sodium in solution which is associated with the active alkali.

$$[Na^+]_{act} = [NaOH] + 2\,[Na_2CO_3] + [NaHCO_3] + [NaHSO_3] + 2\,[Na_2SO_3]$$

If NaOH, $NaHCO_3$, Na_2CO_3, $NaHSO_3$ or Na_2SO_3 solids are added to double alkali solutions, the increase in sodium ion in solution is "active sodium." If Na_2SO_4 or NaCl, for example, is added, the increase in sodium ion is "inactive sodium." Active sodium is not increased by the dissolution of SO_2 in scrubber solutions. Note that the term "active sodium" can be misleading in that the sodium ion doesn't participate in any of the process reactions.

Oxidation

Oxidation in a double alkali system refers to the conversion of TOS to sulfate by one of the following equations:

$$HSO_3^- + 1/2\ O_2 \rightarrow SO_4^{=} + H^+ \qquad (12)$$

$$SO_3^{=} + 1/2\ O_2 \rightarrow SO_4^{=} \qquad (13)$$

Simple oxidation of SO_2 to SO_3 in the flue gas is also considered oxidation in the double alkali system:

$$SO_2 + 1/2\ O_2 \rightarrow SO_3 \qquad (14)$$

Oxidation in the system has the effect of changing active sodium to inactive sodium, or active alkali to inactive alkali.

Oxidation may occur in any part of the system: in the scrubber, the reaction vessels, or in the solids separation equipment. In general, rate of oxidation in the system is thought to be a function of rate of dissolution of oxygen, pH of the scrubbing solution, impurities present in solution and concentration of reactants. Oxidation rate is thus affected by composition of the scrubbing liquor (scrubbing liquors containing high concentrations of dissolved salts may absorb oxygen more slowly), oxygen content of the flue gas, impurities in the coal and lime or limestone, and the design of the equipment (the regeneration and solids separation sections of the system in particular can be designed to limit dissolution of oxygen and number of scrubber contact stages is extremely important).

Oxidation rate is expressed as a percentage and is calculated from an overall material balance on the system:

$$\text{Oxidation rate } (\%) = \frac{[SO_4^{=} \text{ leaving the system (moles)}]}{[\text{Total sulfur collected (moles)}]} \times 100$$

Sulfate leaving the system is total moles of sulfate in the solid waste plus any sulfate in the associated liquor.

Sulfate Regeneration

This term is a misnomer. What is really meant is sulfate removal from the system with regeneration of active alkali from inactive sodium sulfate. (See Sulfate Removal.)

Sulfate Removal

Sulfate is removed from the system with regeneration of active alkali from inactive sodium sulfate. Examples of these sulfate removal reactions are given below:

$$Na_2SO_4 + Ca(OH)_2 + 2H_2O \rightarrow 2NaOH + \underset{\text{(gypsum)}}{CaSO_4 \cdot 2H_2O\downarrow} \tag{15}$$

$$Na_2SO_4 + 2CaSO_3 \cdot 1/2\ H_2O + H_2SO_4 + 3H_2O \rightarrow 2NaHSO_3 + \underset{\text{(gypsum)}}{2CaSO_4 \cdot 2H_2O\downarrow} \tag{16}$$

$$\begin{gathered} y\ Na_2SO_4 + x\ NaHSO_3 + (x+y)\ Ca(OH)_2 + (z-x)\ H_2O \rightarrow \\ (x+2y)\ NaOH + \underset{\text{(mixed crystal or solid solution)}}{x\ CaSO_3 \cdot y\ CaSO_4 \cdot z\ H_2O\downarrow} \end{gathered} \tag{17}$$

$$Na_2SO_4 + 3H_2O \xrightarrow{\text{Electrolytic cell}} 2NaOH + H_2SO_4 + H_2 + 1/2\ O_2 \tag{18}$$

Sulfate removal should be accomplished in an environmentally acceptable manner; a simple purge of soluble Na_2SO_4 from the system to land or waterway disposal is not acceptable in large quantities in most cases.

Softening

This term is used to describe various methods used to lower the dissolved calcium ion concentration in regeneration solutions. The purpose of softening the scrubbing liquor before recycling to the scrubber is to assure that it is subsaturated with respect to gypsum. This reduces the gypsum scaling potential in the scrubber. Following are examples of softening reactions:

$$Ca^{++} + Na_2CO_3 \rightarrow 2Na^{+} + CaCO_3\downarrow \tag{19}$$

$$Ca^{++} + Na_2SO_3 + 1/2\ H_2O \rightarrow 2Na^{+} + CaSO_3 \cdot 1/2\ H_2O\downarrow \tag{20}$$

$$Ca^{++} + CO_2 + H_2O \rightarrow 2H^{+} + CaCO_3\downarrow \tag{21}$$

In each of the above reactions, calcium ions are removed from solution as part of an insoluble material, outside the scrubber system. Reactions (19) and (21) are referred to as carbonate softening. Reaction (20) is considered sulfite softening. Generally, dilute systems employ carbonate softening while concentrated systems resist scaling due to the high sulfite concentrations which prevent high Ca^{++} ion in the scrubber liquor.

Dilute vs. Concentrated System

Dilute or concentrated refers to the active alkali concentration in a particular system. This differentiation is made because, in theory at least, based on their solubility products in water, both $CaSO_4$ and $CaSO_3$ should not precipitate from a solution of sulfite and sulfate simultaneously when using relatively small quantities of lime slurry for regeneration, unless the concentrations of sulfite and sulfate are present in a certain ratio. This can be shown by dividing one solubility product equation by the other:

$$\frac{[Ca^{++}]\ [SO_4^{=}] = Ksp_1}{[Ca^{++}]\ [SO_3^{=}] = Ksp_2}$$

from this, cancelling Ca^{++} ion concentration,

$$\frac{[SO_4^{=}]}{[SO_3^{=}]} = \text{constant}$$

The ratio of sulfate to sulfite for simultaneous precipitation of $CaSO_4$ and $CaSO_3$ is shown to be a constant.

The constant in the above equation is the ratio of solubility product constants of calcium sulfate and sulfite. Then, in theory, if the ratio of sulfate to sulfite is higher than this constant, only calcium sulfate should precipitate; and if the ratio is lower than the constant, only calcium sulfite should precipitate.

This very simplified consideration of the chemistry given above is clouded in the "real world" by factors that contribute to non-ideal behavior in these systems. These factors include changes in ionic activities in solutions containing high electrolyte concentrations and evidence of coprecipitation of calcium sulfite and sulfate in the form of a "mixed crystal" or "solid solution" in a manner which is not completely understood at present.[2,3]

With due consideration to the non-ideal behavior of these systems, however, under given conditions, a ratio of sulfate to sulfite in solution can be determined at which the previously cited examples hold. The ratio establishes a definition for "dilute" or "concentrated" double alkali systems. When the ratio is such that either gypsum or both gypsum and calcium sulfite will precipitate from the solution with the addition of slaked lime, then the system is "dilute."

Lime or Limestone Stoichiometry

Lime or limestone stoichiometry can be expressed as a percentage, based on an overall material balance around the system:

$$\text{Lime Stoichiometry} = \frac{\text{moles CaO added}}{\text{mole sulfur collected}} \times 100$$

$$\text{Limestone Stoichiometry} = \frac{\text{moles CaCO}_3\text{ added}}{\text{mole sulfur collected}} \times 100$$

Lime or limestone stoichiometry is an indication of the efficiency of usage of lime or limestone in the system. Ignoring alkali components in the flyash collected and the alkalinity added with sodium make-up to the system, 100% stoichiometry is complete utilization; stoichiometries over 100% represent less efficient utilization of lime or limestone. Stoichiometries under 100% indicate alkalinity from other sources.

Feed Stoichiometry

This is calculated by a material balance around the scrubber. It is usually expressed as the ratio:

$$\text{Feed Stoich} = \frac{\text{Liquor SO}_2\text{ Capacity (moles/liter)} \times \text{Flow (liters/min)}}{\text{SO}_2\text{ (mole/min)}}$$

This ratio is evaluated for streams entering the scrubber.

Feed stoichiometry is a measure of the ability of the incoming liquor to react with or absorb all of the incoming SO_2 in the scrubber, assuming ideal contact of gas and liquor. Feed stoichiometry above 1.0 is required for high SO_2 removal. At feed stoichiometries at or below 1.0, assuming ideal contact between the gas and liquor, there will be significant equilibrium SO_2 partial pressure above the liquor, and thus SO_2 removal is theoretically limited to the value calculated on the basis of this SO_2 partial pressure in the exiting flue gas.

SIGNIFICANT PROCESS AND DESIGN FACTORS

A commercial double alkali system must be designed to remove the desired quantity of sulfur oxides from a given flue gas stream, while operating in a reliable manner and discharging environmentally acceptable solid waste product. In fulfilling these design objectives, cost is also an important factor.

SO_2 Removal

The fact that small quantities of sulfur dioxide can be removed from large amounts of relatively inert gas by cyclic processes involving absorption into aqueous solutions of sodium sulfite/bisulfite has been known for some time. Johnstone et al.[4] published a paper in 1938 giving data on the vapor pressure of SO_2 over solutions of sulfite/bisulfite and methods of calculating these equilibrium values under various conditions. The equilibrium partial pressure of SO_2 above sulfite/bisulfite solutions, the theoretical limit which a practical design can approach, is generally a function of solution temperature, pH, concentration of sulfite/bisulfite and total ionic strength. Since Johnstone's work a number of organizations have pursued this technology with laboratory, pilot plant and full scale applications for flue gas desulfurization, and many have demonstrated the ability for high removal efficiencies. (It should be noted that although Johnstone's work was aimed at cyclic processes with thermal regeneration, such as the Davy Powergas system, the vapor pressure data is also applicable to double alkali systems which use chemical regeneration.)

Once methods have been established to determine equilibrium SO_2 vapor pressure over scrubbing solutions, of the various concentrations to be encountered in an operating system, it becomes a matter of standard chemical engineering practice to design adequate gas absorption equipment to accomplish the desired SO_2 removal in a system. For comparison, it should be noted that the design of lime/limestone slurry absorption equipment is further complicated by the kinetics of dissolution of the lime or limestone, the particle size of the suspended material, and the crystal morphology of the lime or limestone.

Reliable Operation

System reliability can be adversely affected by two classes of problems, chemical and mechanical.

The mechanical problems include malfunction of instrumentation and mechanical and electrical equipment such as pumps, filters, centrifuges, and valves. These problems in a commercial FGD system can be minimized by careful selection of materials of construction and equipment and by providing spares for certain equipment such as pumps and motors which are expected to be in continuous operation and are prone to failure after a relatively short period of operation. Another important consideration in minimizing mechanical problems is the institution of a good preventive maintenance program.

The chemical (or physical/chemical) problems which may be associated with a double alkali system include scaling, excessive sulfate build-up, production of poor-settling solid waste product, water balance and build-up of non-sulfur solubles which enter the system as impurities in the coal or lime. Each of these factors is associated with reliable system operation, or production of an environmentally acceptable solid waste.

a. Scaling - One of the primary reasons, and probably the most important, for development of double alkali processes was to circumvent the scaling problems associated with lime/limestone wet scrubbing systems. Therefore, a double alkali system should be designed to operate in a non-scaling manner.

Scaling is caused by precipitation of calcium compounds from process liquors, on the surfaces of various components of the system. When this occurs in the scrubber it is particularly troublesome since the flue gas path through the scrubber, if affected, could cause shutdown of the boiler-scrubber system and lower reliability.

Since scrubbing in double alkali systems employs a clear solution rather than a slurry, there is a tendency to ignore potential scaling problems. Testing experience with double alkali systems has indicated, however, that scaling can occur and indeed the problem should be a legitimate concern in the design of any system. Both gypsum and carbonate scale build-up has been recognized in these systems. Gypsum scaling is caused by the reaction of soluble calcium ion with sulfate ion formed in the system through oxidation of the absorbed sulfur dioxide or from absorbed sulfur trioxide according to the reaction:

$$Ca^{++} + SO_4^{=} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O \downarrow \qquad (22)$$

In dilute systems gypsum scaling is controlled by softening the regenerated liquor prior to recycling to the scrubber while in concentrated systems gypsum is not a problem since the high sulfite concentration keeps the Ca^{++} ion low. Softening ensures that the liquor recycled to the scrubber system is unsaturated with respect to gypsum; therefore, with proper softening even if some sulfate is formed in the scrubber, the liquor will not be saturated with gypsum and cause scaling on the inside surfaces of the scrubber. In concentrated active alkali systems, a special softening step is not necessary since high sulfite concentration is maintained throughout the system. This sulfite maintains a low calcium ion concentration (sulfite softening), and thus maintains the scrubbing solution unsaturated with respect to gypsum.

Based on experience gained in lime/limestone scrubbing testing, a certain factor of safety in the prevention of gypsum scaling probably exists in double alkali systems. Gypsum has been found to supersaturate easily to about 130% saturation; thus, even if sulfate formation is higher than expected, gypsum may not precipitate in the scrubber until the liquor is over 130% of saturation with respect to gypsum.

Carbonate scaling usually occurs as a result of localized high pH scrubbing liquor in the scrubber where CO_2 can be absorbed from the flue gas to produce carbonate ion. This ion subsequently reacts with dissolved calcium to precipitate calcium carbonate scale according to the following series of reactions:

Carbon dioxide absorption by high pH liquor:

$$CO_2 + 2\ OH^- \rightarrow CO_3^= + H_2O \qquad (23)$$

Calcium carbonate scaling:

$$CO_3^= + Ca^{++} \rightarrow CaCO_3 \downarrow \qquad (24)$$

Based on experience with the General Motors full scale double alkali system,[5] carbonate scaling could occur with scrubber liquor pH above 9. At lower pH, the carbonate/bicarbonate equilibrium system tends to limit the free carbonate ion and thus prevent precipitation of calcium carbonate:

$$H^+ + CO_3^= \rightleftarrows HCO_3^- \qquad (25)$$

Thus, carbonate scaling can be eliminated by control of pH in the scrubber.

b. Production of Poor-Settling Solids - Under certain conditions, the waste product solids produced in the regeneration sections of various double alkali systems have a tendency not to settle from the scrubber liquors. This creates problems in the operation of settlers, clarifiers, reactor clarifiers, filters and centrifuges. Although the phenomenon has been observed in the laboratory testing conducted by EPA on dilute systems and in the laboratory and pilot plant work conducted by Arthur D. Little, Inc. (ADL) on dilute and concentrated systems, it is not completely understood, but is thought to be a function of reactor kinetics.[3]

Some of the factors thus far identified which appear to affect the solids settling properties are reactor configuration, concentration of soluble sulfate, concentration of soluble magnesium and iron in the liquor, concentration of suspended solids in the reaction zones, and use of lime vs. limestone for reaction. Based on laboratory work in dilute systems (about 0.1 M active sodium) using limestone, it appears that solids settling characteristics degraded significantly at soluble sulfate levels above 0.5 M. Based on laboratory work with concentrated systems (about 0.45 M TOS, 5.4 pH, 0.6 M sulfate) using limestone, marked degradation of solids settling properties occurred at a magnesium level of 120 ppm and virtually no settling of solids occurred at the 2000 ppm magnesium level. Equal degradation of solids settling properties also occurred in concentrated systems when the sulfate level was raised to the 1.0 M level while maintaining low magnesium level (about 20 ppm) and keeping other variables constant.[3]

Envirotech[6] advocates the recycle of precipitated solids from the thickener underflow to the reaction zones in an effort to grow crystals which settle faster and are more easily filtered.

ADL cites reactor configuration as being important in the production of solids with good settling and filtration characteristics.[3] Their basis for this is comparative tests of a simple continuous flow stirred tank reactor (CFSTR) with the ADL/Combustion Equipment Associates (ADL/CEA) designed reactor system under similar conditions. The ADL/CEA reactor system appeared to give better settling solids over a greater range of conditions than a simple CFSTR.

c. Reliability/Availability Data - Three full-scale utility and several industrial boiler double alkali systems have started operation in Japan since 1973. Although detailed availability data is not shown here for these units, few, it any, significant operating problems have been reported for them. Ando has given more detail on these systems in another paper.[18]

In the U.S. the 20 Mw utility prototype system constructed by CEA/ADL for Southern Services at Gulf Power's Scholz station has been in operation for almost a year. That system started up in February 1975 and operated intermittently through January 1976. It was shut down for modification and repairs from mid-July through mid-September 1975. Excluding that 2-month period, the system maintained a 90% availability during the first, approximately 1 year of operation. Most of the problems experienced were related to the filtration system.

FMC's double alkali kiln control system has been in operation at their Modesto, California, plant for over 5 years. That system is equivalent to 10 Mw based on gas rate or about 30 Mw based on the size of the regeneration system and SO_2 absorbed. As of approximately a year ago, that system reportedly was available for control of SO_x emissions from the barium reduction kilns 100% of the time required. The kilns were in service 95% of the time with scheduled shutdowns for routine maintenance and/or change of chemical process service every 3 or 4 months.

Environmentally Acceptable Solid Waste

Double alkali systems should be designed to produce an environmentally acceptable end product. Some of the properties which can be ascribed to such a solid waste product are listed below:

- Non toxic
- Low soluble solids content, non-leachate
- Low moisture content
- Non-thixotropic
- High compressive or bearing strength

In order to generate waste product solids which have properties approaching those listed, the following design considerations are appropriate:

- Sulfate removal
- Water balance/waste product (cake) wash
- Gypsum vs. calcium sulfite
- Sludge fixation technology

a. Sulfate Removal - In double alkali systems, some of the sulfur removed from the flue gas takes the form of soluble sodium sulfate due to oxidation in the system, thus changing some of the active sodium to the inactive variety. When sodium in the system is converted to the inactive form (Na_2SO_4), it is relatively difficult to convert back to active sodium. To convert inactive sodium to active sodium, sulfate ion must be removed from the system in some manner, while leaving the sodium in solution. The alternative to this is to remove the sodium sulfate from the system at the rate it is being formed in the system. This alternative is not desirable since it is wasteful of sodium and generally is carried out by allowing the sodium sulfate to be purged from the system in the liquor which is occluded in the wet solid waste product.[7] The solid waste product can then potentially contribute to water pollution due to leachability. Water run-off can lead to contamination of surface water, while leaching and percolation of the leachate into the soil can result in contamination of the ground water in the vicinity of the disposal site. Failure to allow for sulfate removal from double alkali systems will ultimately result in a) precipitation of sodium sulfate somewhere in the system if active sodium is made up to the system, or b) in the absence of make-up, eventual deterioration of the SO_2 removal capability due to the loss of active sodium from the system.

Equations (15), (16), (17) and (18), shown previously under the definition of "sulfate removal", describe several sulfate removal techniques which have been used in FGD system pilot tests.

The first equation depicts the sulfate removal technique used in dilute active alkali systems:

$$Na_2SO_4 + Ca(OH)_2 + 2H_2O \rightleftarrows 2NaOH + \underset{\text{(gypsum)}}{CaSO_4 \cdot 2H_2O\downarrow} \qquad (15)$$

Concerning the full-scale dilute alkali system installed and operating at the Parma, Ohio, transmission plant of General Motors, and dilute systems in general, Phillips[5] stated:

> "The presence of Na_2SO_4 in the scrubber effluent is the prime factor influencing the design of the regeneration system. Na_2SO_4 is not easily regenerable into NaOH using lime. The reason being that the product, gypsum, is relatively soluble. . . . Na_2SO_4 cannot be causticized in the presence of appreciable amounts of $SO_3^=$ or OH^- because Ca^{++} levels are held below the $CaSO_4$ solubility product. To provide for sulfate causticization, the system must be operated at dilute OH^- concentrations below 0.14 molar. At the same time, $SO_4^=$ levels must be maintained in the system at sufficient levels to effect gypsum precipitation. . . . We selected 0.1 molar OH^- and 0.5 molar $SO_4^=$ as a design criteria."

In a previous paper, Phillips[8] showed a plot of equilibrium caustic formation in $Ca(OH)_2$-Na_2SO_4 solutions at 120°F which is the basis for selection of the design criteria. The essence of this discussion is that if the active sodium concentration is sufficiently dilute, sulfate can be removed from the system by simple precipitation as gypsum by reaction of lime with sodium sulfate.

Since, as explained above, this reaction will not proceed to a great extent in concentrated active alkali systems, other techniques must be employed to effect sulfate removal in these systems.

The second equation depicts a technique which is used in the full scale double alkali systems in Japan, and which has been pilot tested by ADL under contract with EPA:

$$Na_2SO_4 + 2CaSO_3\cdot 1/2\ H_2O + H_2SO_4 + 3H_2O \rightarrow 2NaHSO_3 + \underset{\text{(gypsum)}}{2CaSO_4\cdot 2H_2O\downarrow} \qquad (16)$$

This technique is used to precipitate gypsum by dissolving calcium sulfite in acidic solution thus increasing the Ca^{++} in solution enough to exceed the solubility product of gypsum. Ideally according to equation (16) 2 moles of gypsum should be precipitated for each mole of sulfuric acid added. In practice, however, this is not the case since any material which functions as a base can consume sulfuric acid and reduce the efficiency of this reaction for its intended purpose.[3] Unreacted lime or limestone, sulfite ion and even sulfate ion can consume sulfuric acid thus lowering sulfate removal from the system.

Conceivably, this method of sulfate removal may be economically unattractive in applications with very high oxidation rates, and where the gypsum produced must be discarded. The economic picture is considerably changed where this system is used merely as a slip-stream treatment to supplement other sulfate removal methods and/or where the solid product gypsum is saleable as is the case in Japan.

The third equation describes a phenomenon which has been referred to as mixed crystal or solid solution formation:

$$\begin{aligned} &x\ NaHSO_3 + y\ Na_2SO_4 + (x+y)\ Ca(OH)_2 + (z-x)\ H_2O \rightarrow \\ &\quad (x+2y)\ NaOH + x\ CaSO_3 \cdot y\ CaSO_4 \cdot z\ H_2O \downarrow \\ &\qquad \text{(mixed crystal or solid solution)} \end{aligned} \tag{17}$$

This phenomenon is described by R.H. Borgwardt of EPA[2] as it applies to lime/limestone wet scrubbing based on pilot plant investigations. A similar phenomenon has been observed by ADL in some of their early pilot testing of double alkali systems in conjunction with CEA, and later in the EPA/ADL dual alkali test program.

It appears that under certain conditions the solids precipitated in lime/limestone and double alkali systems contain sulfate, sulfite and calcium; however, the liquor from which these solids precipitated, appears to be subsaturated with respect to gypsum. This is based on the fact that pure gypsum crystal could be dissolved in the mother liquor from which the mixed crystal/solid solution was precipitated. In addition, the solid material was examined by X-ray diffraction and found to contain no gypsum; infrared analysis confirmed the presence of sulfate.

Borgwardt found that the molar ratio of sulfate to sulfite in these solids was primarily a direct function of sulfate ion activity in the mother liquor. In pilot test work with lime/limestone scrubbing, with little or no chlorides present and normal magnesium level (below 1000 ppm) in solution, the sulfate to sulfite molar ratio in the mixed crystal solids was found to reach a maximum level of 0.23. This is equivalent to a $[SO_4^{=}]$/total $[SO_x]$ ratio in the solids of 0.19.

In pilot test work with concentrated double alkali systems, ADL observed the simultaneous precipitation of sulfate and sulfite with calcium in lime and limestone treatment of concentrated double alkali scrubbing liquors. This phenomenon was surprising at first, in light of the reasoning which led to the development of dilute double alkali systems; i.e., gypsum cannot be precipitated from solutions containing high active alkali concentrations. It was a simple technique for sulfate removal in concentrated systems. The $[SO_4]$/ total $[SO_x]$ ratio observed in pilot double alkali work was as high as 0.20.[3] Coincidentally, this is the same value observed by Borgwardt in lime/limestone testing. This leads to the suspicion that the same phenomenon is occurring in both processes. The mother liquor from which these solids were precipitated was also found to be subsaturated in gypsum, and when the solids were examined, pure gypsum was not found.

Based on the observed data, it appears reasonable to design a concentrated active alkali system for a particular situation in which the system oxidation rate is below about 20%.[3] In this case, sulfate can be removed at the desired rate, without the necessity for purging Na_2SO_4 or supplementing the system with other complex methods of sulfate removal.

The fourth equation shows sulfate removal as sulfuric acid in an electrolytic cell:

$$Na_2SO_4 + 3H_2O \xrightarrow[\text{cell}]{\text{Electrolytic}} 2NaOH + H_2SO_4 + H_2 + 1/2\ O_2 \quad (18)$$

This method is the basis for operation of the Stone and Webster/ Ionics process sulfate removal technique. In Japan, Kureha/Kawasaki has pilot tested the Yuasa/Ionics electrolytic process for sulfate removal in conjunction with their double alkali process. They feel that this process will be less expensive overall than the presently used sulfuric acid addition method. In addition, they feel that sodium losses from the system can be cut in half through the use of this method: from 0.018 moles Na loss/mole SO_2 absorbed to 0.009 moles Na loss/mole SO_2 absorbed.

Another approach to sulfate control is to limit oxidation. With sufficient limitation of oxidation, by process and equipment design, it may be possible to control sulfate by a small unavoidable purge of Na_2SO_4 with the solid waste product. To design for minimum oxidation, there should be minimum residence times in equipment where the scrubber liquor is in contact with oxygen-containing flue gas, and all reactors, mixers, and solids separation equipment should be designed to minimize absorption of oxygen from air. In addition, it has been reported[9] that oxidation of scrubber liquors can be minimized by maintaining very high ionic strength. One possible explanation for this is that high ionic strength liquors are poor oxygen absorbers and that oxidation in these systems is oxygen absorption rate limited.

b. Water Balance and Waste Product (Cake) Washing - In order to operate a closed system to avoid potential water pollution problems, system water balance is a primary concern. Water cannot be added to the system at a rate greater than the normal water losses from the system.

Generally fresh water is added to a D/A system to serve many purposes. These include:

- Saturation of flue gas
- Pump seal needs
- Demister washing needs
- Slurry make-up needs
- Waste product washing
- Tank evaporation

On the other hand water should only leave the system in the following ways:

- Evaporation by the hot flue gas
- Water occluded with solid waste product
- Water of crystallization in solid waste product

Careful water management, part of which is the use of recycled rather than fresh water wherever possible, is necessary in order to operate a closed system.

As previously indicated, disposal of wet solid waste containing soluble salts is ecologically undesirable. In addition, allowing active alkali or sodium salts to escape from the system is an operating cost factor. Sodium make-up to double alkali systems is usually accomplished by adding soda ash (recently quoted at $70 per ton) at some point in the system. Thus, both ecological and economic considerations dictate that waste product washing is desirable.

A rotary drum filter, belt filter or centrifuge is usually the equipment in which the final solids separation is made. This equipment can be designed to permit solids washing with fresh water.

One concern in waste product washing is the extent to which the cake should be washed. At present, there are no federal regulations concerning the amount of total dissolved solids (TDS) which is permitted in a waste product which is to be disposed of as landfill. The future, however, may unfold stringent regulations in this area. One obvious consideration in waste product washing is system water balance. Unlimited waste product washing is not possible if a closed system operation with no liquid stream discharge is a goal. Another more subtle reason for limiting waste product washing is the potential problem of non-sulfur/calcium solubles build-up in the system. These non-sulfur/calcium solubles enter the system with the fly ash, flue gas, the lime and/or limestone and the make-up water. Of these, probably the soluble material in highest concentration would be sodium chloride which results from the absorption of HCl from the flue gas by the scrubber solution. A material balance around the system at steady state necessitates that solubles leave the system at the rate they enter. Thus, depending upon how well the waste product is washed, a certain level of non-sulfur solubles will be established in the system. Since the only mechanism for these solids to leave the system is as part of the wet solid waste, a certain purge rate is necessary. This purge also necessitates the loss of some sodium from the system. Practical limitations in filter design and water balance probably would limit a system to two or three "displacement washes" of the waste product (one displacement wash means washing with an amount of fresh water equivalent to the amount of water contained in the final wet waste product per unit of waste). Depending on the characteristics of the waste product and the design of the washing system, one displacement wash can reduce the solubles content of the waste product by as much as 80%.

c. Gypsum vs. Calcium Sulfite - Although gypsum ($CaSO_4 \cdot 2H_2O$) is more soluble than calcium sulfite hemihydrate ($CaSO_3 \cdot 1/2\ H_2O$), gypsum may be considered a more environmentally acceptable end product. The solubility of gypsum in water is about 0.25%; that of calcium sulfite is on the order of 0.0025%. It is interesting to note that while gypsum is a naturally occurring mineral, calcium sulfite is not found in nature. In the solid waste product (sludge) from lime/limestone and double alkali FGD systems, gypsum has better handling properties than calcium sulfite. Sludges containing a high ratio of gypsum to calcium sulfite are less thixotropic, better settling, more easily filtered, and can be more completely dewatered than sludges containing a high proportion of calcium sulfite. Another important characteristic which has been attributed to high gypsum (as opposed to calcium sulfite) sludges is their higher compressive or bearing strength. Typically, lime/limestone systems which generate solids having a high proportion of calcium sulfite can be filtered to 40-50% solids, while some double alkali systems producing high gypsum cake can be filtered to over 65% solids.[10]

Some explanation for the behavior of these sludges is given by Selmeczi and Knight.[11] Although filter cakes appear dry, they still contain a considerable amount of water and thus, upon vibration or application of stress, they have the tendency of again becoming fluid. This thixotropic property and high moisture content are both explained by the morphology of calcium sulfite clusters. Due to the highly open, porous or sponge-like nature of these clusters, a considerable amount of water is retained in the clusters. The calcium sulfite crystals are rather fragile and break under pressure releasing some of the water, which results in the sludge becoming fluid.

It is possible in some double alkali systems to produce a gypsum product. In Japan, where by-product gypsum is a saleable product, the calcium sulfite solids produced are oxidized completely to gypsum in a separate oxidation process tacked on to the tail end of the system. In applications where high excess combustion air is present in the boiler or where low sulfur coal is burned or a combination of these conditions is present, the oxidation rate in the system tends to be high (possibly about 90%) and the proportion of gypsum in the sludge tends to be high. In some dilute systems the proportion of gypsum in the sludge can be increased by augmental aeration of the scrubbing liquor.[12] Crystal seeding techniques used in conjunction with augmental aeration can produce relatively coarse grained gypsum crystals with good dewatering and structural properties in the final waste product.

d. Sludge Fixation Technology - Chemical or physical fixation of the sludge produced in a double alkali system is another potentially important means of producing an environmentally acceptable solid waste product. This technology is under investigation by I.U. Conversion Systems, Inc., Chicago Fly Ash, Dravo Corporation, and Chemfix Corporation. Most of their efforts are concentrated on sludge produced from the more prevalent lime/limestone systems; however, there is also some evaluation of double alkali sludges. The objective of sludge fixation technology is the production of a non-toxic, unleachable solid waste product which has reasonably high load bearing strength. If double alkali sludges are amenable to this type of treatment, the need to reduce soluble sulfates in the solid waste product becomes less of a problem. Some sodium sulfate has been found to be physically or chemically tied up in the solid calcium sulfate/sulfite crystal lattice;[3] however, the extent of this phenomenon is not generally considered to be adequate to remove all of the sodium sulfate produced via oxidation. Sludge fixation technology may be a mechanism by which additional sodium sulfate can be removed from the system without adverse environmental effects. There is some concern as to whether this is viable, however, since sludge fixation chemistry involves pozzolanic reactions between calcium compounds and flyash components in the sludge which may only involve multivalent ions rather than monovalent sodium. In other words, monovalent ions such as sodium may either a) not take part in the pozzolanic reactions, or b) inhibit or limit such reactions. Further investigation is called for in this area.

Dilute vs. Concentrated System

The selection of a dilute or concentrated double alkali system is an important design consideration in any application. In general it can be stated that the concentrated systems are more suited to applications in which oxidation is expected to be relatively low, and that, conversely, dilute systems are favored in applications where oxidation rates are high. High sulfur Eastern coal applications on utility boilers where excess air is controlled carefully and maintained at the lowest value consistent with complete combustion, the concentrated systems may be favored. On the other hand, in utility or industrial boiler applications where Western low sulfur coal is burned, and/or where control of oxidation is difficult due to high excess air, the dilute systems may be favored.

Oxidation rate is promoted when low sulfur coal is burned, since the ratio of oxygen to sulfur dioxide in the flue gas is higher than in high sulfur coal applications. Since oxidation is a strong function of the rate of absorption of oxygen, liquor which is dilute in TOS is subject to having a greater proportion of these species oxidized by a given amount of absorbed oxygen than one in which the TOS is more concentrated.

Under a given set of conditons without consideration given to waste disposal, a concentrated system can be installed at lower capital cost than a dilute system as previously discussed; however, the desirability to produce a manageable solid waste (dilute systems can be designed to produce high gypsum sludges) could, in cases, override the capital cost issue.

Sodium Consumption

Sodium consumption is an important performance criterion for a double alkali system not so much from the viewpoint of economics as from the viewpoint of producing a leachable waste product as stressed under sulfate removal. Sodium consumption is only a minor factor in the operating cost of a system representing about 2% of the annual operating cost. Thus even if soda ash make-up to a system were to increase by 100% over the expected value, operating cost might only be increased by a factor of 1.02. On the other hand the environmental consequences of higher sodium consumption may be significant if the sodium is leachable from the waste product to the environment.

A logical way to measure sodium consumption is in moles of Na consumed per mole of sulfur removed by the system. A value of 0.05 moles of Na make-up per mole of sulfur removed (equivalent to 0.025 moles of Na_2CO_3 make-up per mole of sulfur removed) appears to be a reasonable design target based on present U.S. technology. This target is achievable in concentrated alkali systems burning relatively high sulfur coal (over 3% sulfur) and having a relatively low oxygen content flue gas. In Japan sodium make-up is reportedly as low as 0.02 moles Na/mole sulfur removed for some systems.

Calcium Consumption

A logical way to specify calcium consumption is as calcium stoichiometry moles of calcium added per mole of sulfur removed (or collected). A calcium consumption of 0.98 to 1.0 appears to be a reasonable design target for concentrated double alkali systems (values under 1.0 are possible due to the additional alkali added with sodium make-up to the system).

Power Consumption

Design targets for power consumption can be in the range of 1-2% of power plant output without reheat or under 3% even with 50°F of reheat. These figures are based on a system which has a scrubber pressure drop in the range of 6-8 in. H_2O and a scrubber system L/G ratio of about 10 gal./1000 ft^3. This assumes that the power plant is equipped with some means of efficient particulate collection upstream of the FGD (e.g., an electrostatic precipitator).

Low power consumption is a major selling point for double alkali FGD systems. Some other FGD systems are estimated to run as high as 10% in power consumption.

Economics

It appears that double alkali systems are now economically competitive with the "first generation" wet alkali lime/limestone slurry scrubbing systems. This is especially true in cases where the lime or limestone system would be required to be equipped with solids separation equipment (i.e. thickener and filter).

Some of the factors that allow a double alkali system to be less expensive than a lime/limestone system both in initial cost and annualized operating cost are:

- Lower scrubber liquid/gas ratio (L/G)
- Lower scrubber pressure drop (ΔP)
- Simpler scrubber design
  - fewer stages
  - no slurry in scrubber
- Less exotic materials of construction
- Solid waste with better handling properties

It is estimated that a new (or simple retrofit) double alkali system could be installed at a 200 Mw or larger plant for a cost of $50 - 60/kw and operated at 2.5 - 3 mills/kw-hr excluding sludge disposal (which would vary with the particular application). These estimates are based on the following assumptions:

200 Mw or larger system
3 - 4% sulfur coal
80% load factor
Capital charges @15-16% of capital investment, annually
Maintenance @3-4% of capital investment, annually
Operation with two operators/shift
Power @$0.02/kw-hr
Soda Ash @ $65-70/ton
Lime @ $30-33/ton

It appears that economics of scale for units above 200 Mw in size are not very significant since larger units would probably involve the installation of additional modules. There would be some economy of scale in the savings on design, engineering, site preparation, etc.

THE EPA DOUBLE ALKALI RESEARCH, DEVELOPMENT AND DEMONSTRATION PROGRAM

EPA has been involved in the development of double alkali technology since the Second International Lime/Limestone Wet-Scrubbing Symposium held in New Orleans in November 1971. Some of the incentive to develop this technology stemmed from a paper presented by R.J. Phillips[8] of General Motors (GM) concerning GM's laboratory and pilot plant work with a dilute double alkali system which appeared very encouraging at that time in light of the difficulties which were being experienced with lime/limestone systems. The development of double alkali technology by EPA has followed an orderly progression of scale from laboratory to pilot plant to prototype and finally to a planned full-scale utility demonstration of the process. Along the way, EPA also embarked on a program to evaluate a full-scale dilute double alkali system in operation at a GM industrial boiler system (32 Mw equivalent).

Some initial laboratory work on regeneration chemistry was done in the EPA laboratories at Research Triangle Park in addition to an initial feasibility study[13] which indicated that double alkali systems might be somewhat lower in capital and operating costs than lime or limestone systems under certain circumstances.

After these initial studies, EPA contracted with Arthur D. Little (ADL) to conduct a study of the double alkali process. The scope of work included in the initial laboratory and pilot plant program was subsequently expanded to include prototype testing at the Scholz Plant of Gulf Power Company where a 20 Mw FGD prototype system was constructed by Combustion Equipment Associates (CEA)/ADL for Southern Services. The construction prototype unit was funded by The Southern Company.

The initial goals of the EPA double alkali program were to:

- Demonstrate reliable system operation
- Demonstrate high SO_2 removal, 95% desirable
- Demonstrate environmentally acceptable sulfate removal schemes
- Minimize soluble material in disposable waste
- Minimize moisture in disposable waste
- Demonstrate closed-loop operation
- Minimize costs
- Minimize Ca^{++} concentration in the scrubber

To date a high degree of success with these goals has been achieved up to the prototype level.

Laboratory

Some of the areas of investigation in the laboratory program are listed below:

- Regeneration of simulated scrubber effluents with lime and limestone
- Sulfate removal by precipitation of a mixed crystal containing $CaSO_4$ and $CaSO_3$ with water of hydration
- Sulfate removal by reaction of Na_2SO_4, $CaSO_3$ and sulfuric acid
- Feasible ranges of sulfate, chloride, magnesium and iron in solution
- Settling characteristics of the product solids
- Fixation of double alkali product solids
- Density, compactability, leachability and permeability of fixed and unfixed solids

Pilot Plant

In the pilot plant both short term and long term runs (∿5 weeks/run) have been conducted to examine various modes of double alkali operation. Some of these modes of operation are listed below:

- Concentrated alkali, sulfuric acid treatment, lime
- Concentrated alkali, two-stage reactor system, lime
- Concentrated alkali, single CSTR reactor, lime
- Dilute alkali w/sulfite oxidation, lime
- Concentrated alkali, multi stage reactor, limestone regeneration
- Extra high concentrated alkali, two stage reactor system, lime

Much of the laboratory and pilot plant work has been reported by LaMantia, et al.[3] in a paper presented at the last EPA Symposium on FGD in Atlanta.

Prototype

The results of testing to date at the 20 Mw prototype system at Gulf Power will be reported in a paper entitled, "Operating Experience -- CEA/ADL Dual Alkali Prototype System at Gulf Power/Southern Services, Inc." by LaMantia, et al.[17]

GM Industrial Boiler

The GM test program at the 32 Mw equivalent system at the Chevrolet Transmission plant in Parma, Ohio is being conducted under an agreement between EPA and GM. The test program design and some chemical analyses are being done for GM and EPA under contract by ADL. The FGD system consists of four Koch tray stainless steel scrubbers, 32 Mw equivalent, and a 40 Mw equivalent regeneration system consisting basically of two mix tanks and two reactor clarifiers. The boiler system consists of two 60,000 lb/hr (of steam) and two 100,000 lb/hr boilers. The GM system is a dilute alkali system.

The system has been operated intermittently since start-up in February 1974. Availability data for the system is not easily defined since there are four boilers and four scrubbers while the steam demand is frequently less than the capacity of one or two boilers. Thus, when there is a problem with one of the systems it can be taken out of service and replaced by one of the stand-by units.

The GM system cost approximately $3.5 million.

A good account of this system was presented in the last EPA Symposium of FGD in Atlanta by Dingo and Piasecki.[16]

Full Scale Utility Demonstration

The RFP by EPA for a full-scale utility demonstration was issued in May 1975. Proposals were received in response in August 1975. At the time of this writing (Feb. 1976), the proposals are in the final stages of evaluation.

The RFP called for a 100 Mw or larger demonstration of the technology. The three proposals receiving final consideration range in size from 192 to 575 Mw. The contract will be let for a four-phase program consisting of:

(I) Process design and cost estimate
(II) Engineering design, construction and mechanical testing
(III) Start-up and acceptance testing
(IV) One year operation and long term testing

A summary of the design criteria requested for the demonstration unit is given below:

- Unit will meet all applicable pollution control regulations
- Pulverized coal fired boiler burning 2.5 - 4.5% sulfur coal
- Instrumentation must be provided to allow accurate material and energy balances at the demonstration plant
- SO_2 control to below 200 ppm emissions by the scrubber
- Stack gas reheat required
- Chemical make-up requirements requested
  - Na 0.05 moles/mole S
  - Ca 1.2 moles/mole S
- Acceptable long-range waste disposal plan

We anticipate conducting an extensive test program at the full-scale facility which would characterize effluents, evaluate chemical consumption, evaluate sludge disposal, optimize operating conditions and evaluate system reliability and process economics.

SUMMARY OF STATUS OF TECHNOLOGY

Based on the initial performance and reliability demonstrated in various double alkali pilot and prototype plants, and in an increasing number of full scale applications in the U.S. and Japan, it appears that this technology has become a viable technological alternative throwaway process to the lime/limestone processes.

Performance of double alkali systems with respect to SO_2 removal is well established. Over 99% SO_2 removal has been achieved with lower than 10 ppm SO_2 concentration in the treated flue gas.

Potential environmental problems associated with waste disposal from these systems may occur due to the presence of soluble sodium salts which could cause water pollution problems in the surface and ground water in the vicinity of disposal sites. A number of techniques to reduce the soluble sodium salts in the waste product have been tested; however, it appears that there will inevitably be a higher level of soluble salts in the waste product from double alkali systems than from lime/limestone systems. With present U.S. technology it appears that the incremental amount of solubles in a double alkali waste product (over that typically present in lime/limestone waste product) is typically about 1-2% on a dry solids basis. Development of sludge fixation technology to change the wet solid waste product to a hard unleachable solid could conceivably reduce this potential problem.

Actual costs for full scale utility boiler applications are not available. Estimates for these applications based on the best available information put double alkali system capital costs in the range of lime/limestone system costs, at about $50-60/kw capital and 2.5-3 mills per kw-hr operating cost.

In the U.S., development has stressed the use of lime rather than limestone as the calcium source in all of the full scale industrial applications and in most of the pilot plant testing in both dilute and concentrated systems. In Japan, limestone rather than lime is used for concentrated systems in full scale utility boiler applications apparently for the operating cost benefit which is derived through the use of the less expensive regenerant. Also, due to differences in the economies of the two countries and state-of-the-art of technology, production of by-product gypsum from the Japanese double alkali systems is prevalent;

whereas, production of a throwaway solid waste product is the general rule with development of the technology in the U.S.

The table presented below shows a total of approximately 3900 Mw equivalent operating and planned double alkali systems in industrial and utility applications in the United States and Japan representing 23 applications of this technology. Of these, approximately 1750 Mw equivalent representing 16 applications are operational. In the U.S. six operational units approximately equivalent to 140 Mw are listed. As a standard for comparison the latest PEDCo Survey of FGD systems in the U.S.[19] indicates a total of 42,160 Mw representing 109 units of operating and planned FGD systems. Of these 3828 Mw representing 22 units are operational.

The development of double alkali technology has obviously stemmed (and benefitted) from both lime/limestone and other soluble sodium scrubbing technology development. Possibly as a result of this and certain inherent advantages of soluble alkali scrubbing, it appears that the reliability established at this point in the development of double alkali technology is greater than that which had been established for the lime/limestone systems at a corresponding stage of development.

SUMMARY OF SIGNIFICANT OPERATING AND PLANNED FULL SCALE DOUBLE ALKALI SYSTEMS

System Operator	System Application	Vendor or Developer	Size (Mw. Equivalent)	Active Alkali	Calcium Sources	Start-Up Date[a]
FMC Modesto Calif. Plant	Reduction kilns	FMC	10 (Gas Rate) 30 (Regen.)	Conc.	Lime	Dec. 1971
Showa Denko KK Chiba, Japan	Oil-fired elec. power boiler	Showa Denko	150	Conc.	Limestone	June 1973
Tohoku Electric Shinsendai Sta., Japan	Oil-fired utility boiler	Kawasaki/ Kureha	150	Conc.	Limestone	Jan. 1974
General Motors Parma, Ohio Plant	4 coal-fired industrial boilers	General Motors	40 (Regen) 32 (Gas Rate)	Dilute	Lime	Mar. 1974
Caterpillar Tractor Co. Joliet, Illinois	2 coal-fired industrial boilers	Zurn Industries	20-30	Dilute	Lime	Oct. 1974
Showa Pet. Chem. Kawasaki, Japan	Oil-fired industrial boiler	Showa Denko	62	Conc.	Limestone	1974
Kanegafuchi Chem. Takasago, Japan	Oil-fired industrial boiler	Showa Denko	93	Conc.	Limestone	1974
Poly Plastic Fuji, Japan	Oil-fired industrial boiler	Showa Denko/Ebara	65	Conc.	Limestone	1974
Kyowa Pet. Chem. Yokkaichi, Japan	Oil-fired industrial boiler	Showa Denko/Ebara	46	Conc.	Limestone	1974
Kinuura Utility Nagoya, Japan	Oil-fired industrial boiler	Tsukishima	63	Conc.	Lime	1974
Daishuwa Paper Fuji, Japan	Oil-fired boiler	Tsukishima	85	Conc.	Lime	1974

SUMMARY OF SIGNIFICANT OPERATING AND PLANNED FULL SCALE DOUBLE ALKALI SYSTEMS (Continued)

System Operator	System Application	Vendor or Developer	Size (Mw, Equivalent)	Active Alkali	Calcium Sources	Start-Up Date[a]
Firestone Pottstown, Pa.	Coal & oil-fired industrial boiler-demonstration	FMC	3	Conc.	Lime	Jan. 1975
Gulf Power Company Scholz Plant (Southern Services)	Coal-fired utility boiler-prototype	A.D. Little/ Combust. Equip. Associates	20	Conc.	Lime	Feb. 1975
Caterpillar Tractor Co. Mossville, Illinois	Coal-fired industrial boiler system	FMC	45-55	Conc.	Lime	Oct. 1975
Sikoku Electric Power Anan, Japan	Oil-fired utility boiler	Kawasaki/ Kureha	450	Conc.	Limestone	1975
Sikoku Electric Power Sakaide, Japan	Oil-fired utility boiler	Kawasaki/ Kureha	450	Conc.	Limestone	1975
Kyushu Electric Power Buzen, Japan	2 oil-fired utility boilers	Kawasaki/ Kureha	(2-450) 900	Conc.	Limestone	(May 1977)
Caterpillar Tractor East Peoria, Illinois	Coal-fired industrial boiler	FMC	100	Conc.	Lime	(1977)
Central Illinois Public[b] Service - Newton #1	Coal-fired utility	Envirotech	575	Conc.	Lime	(1977-78)
Louisville Gas & Electric[b] Cane Run #6 Louisville, Kentucky	Coal-fired utility	CEA/ADL	280	Conc.	Lime	(1977-78)
Springfield Water Light[b] and Power - Dallman #3 Springfield, Illinois	Coal-fired utility	FMC	192	Conc.	Lime	(1977-78)
Caterpillar Tractor Mapleton, Illinois	Coal-fired industrial boiler	FMC	100	Conc.	Lime	(1978)

---

[a] Dates in parentheses are projected start-up dates

[b] These units were proposed in response to the RFP for a full scale utility demonstration

BIBLIOGRAPHY

1. Ponder, W. H., "Status of Flue Gas Desulfurization Technology For Power Plant Pollution Control." Presented at Thermal Power Conference, Washington State University, Pullman, Washington, October 4, 1974.

2. Epstein, M., R. Borgwardt, et al., "Preliminary Report of Test Results from the EPA Alkali Scrubbing Test Facilities at the TVA Shawnee Power Plant and at Research Triangle Park." Presented at Public Briefing, Research Triangle Park, North Carolina, December 19, 1973.

3. LaMantia, C., et al., "EPA-ADL Dual Alkali Program Interim Results." Presented at EPA Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.

4. Johnstone, H. F., et al., "Recovery of Sulfur Dioxide from Waste Gases." Ind. & Eng. Chem., Vol. 30, No. 1, January 1938, pp 101-109.

5. Phillips, R. J., "Operating Experiences with a Commercial Dual-Alkali SO_2 Removal System." Presented at the 67th Annual Meeting of the Air Pollution Control Association, Denver, Colorado, June 9-13, 1974.

6. Cornell, C. F. and D. A. Dahlstrom, "Performance Results on a 2500 ACFM Double Alkali Plant for SO_2 Removal." Presented at the 66th Annual Meeting of A.I.Ch.E., Philadelphia, Pennsylvania, November 11-15, 1973. Condensed version of the paper appeared in December 1973 CEP.

7. Kaplan, N., "An EPA Overview of Sodium-Based Double Alkali Processes - Part II Status of Technology and Description of Attractive Schemes." Presented at the EPA Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14-17, 1973.

8. Phillips, R. J., "Sulfur Dioxide Emission Control for Industrial Power Plants." Presented at the Second International Lime/Limestone Wet-Scrubbing Symposium, New Orleans, Louisiana, November 8-12, 1971.

9. Brady, J. D., "Sulfur Dioxide Removal Using Soluble Sulfites." Presented at Rocky Mountain States Section Air Pollution Control Association, Colorado Springs, Colorado, April 30, 1974.

10. Cornell, C. F., "Liquid-Solids Separation in Air Pollutant Removal Systems." Presented at the ASCE Annual and National Environmental Engineering Convention, Kansas City, Missouri, October 21-25, 1974.

11. Selmeczi, J. G. and R. G. Knight, "Properties of Power Plant Waste Sludges." Presented at the Third International Ash Utilization Symposium, Pittsburgh, Pennsylvania, March 13-14, 1973.

12. Ellison, W., et al., "System Reliability and Environmental Impact of SO_2 Scrubbing Processes." Presented at Coal and The Environment, Technical Conference, Louisville, Kentucky, October 22-24, 1974.

13. Rochelle, G. T., "Economics of Flue Gas Desulfurization." Presented at EPA Flue Gas Desulfurization Symposium, New Orleans, Louisiana, May 14-17, 1973.

14. "Sulfur Dioxide and Flyash Control", FMC Corporation Technical Bulletin. FMC Corporation, Air Pollution Control Operation, 751 Roosevelt Road, Suite 305, Glen Ellyn, Illinois 60137.

15. McGlamery, G. G. and R. L. Torstrick, "Cost Comparisons of Flue Gas Desulfurization Systems." Presented at the EPA Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.

16. Dingo, T. and E. Piasecki, "General Motor's Operating Experience with a Full-Scale Double Alkali Process." Presented at the EPA Symposium on Flue Gas Desulfurization, Atlanta, Georgia, November 4-7, 1974.

17. LaMantia, C.R., R.R. Lunt, R.E. Rush, T. Frank, N. Kaplan, "Operating Experience -- CEA/ADL Dual Alkali Prototype System at Gulf Power/Southern Services, Inc." Presented at the EPA Symposium on Flue Gas Desulfurization, New Orleans, Louisiana, March 8-11, 1976.

18. Ando, J., "Status of Flue Gas Desulfurization and Simultaneous Removal of SO_2 and NO_x in Japan." Presented at the EPA Symposium on Flue Gas Desulfurization, New Orleans, Louisiana, March 8-11, 1976.

19. Summary Report. Flue Gas Desulfurization Systems. Nov. - Dec. 1975 Prepared by PEDCo Environmental Specialists for U.S. EPA under Contract No. 68-02-1321.

OPERATING EXPERIENCE--CEA/ADL DUAL ALKALI PROTOTYPE SYSTEM AT GULF POWER/SOUTHERN SERVICES, INC.

Charles R. LaMantia and Richard R. Lunt

Arthur D. Little, Inc.
Cambridge, Massachusetts

Randall E. Rush

Southern Services, Inc.
Birmingham, Alabama

Thomas M. Frank

Combustion Equipment Associates, Inc.
New York, New York

Norman Kaplan

Industrial Environmental Research Laboratory
Environmental Protection Agency
Research Triangle Park, North Carolina 27711

ABSTRACT

As part of the Dual Alkali Program being conducted by Arthur D. Little, Inc. for EPA's Industrial Environmental Research Laboratory-RTP, a one-year test is being conducted on the 20-megawatt dual alkali SO_2 control process at Gulf Power Company's Scholz Steam Plant in Sneads, Florida. The system was developed, designed and installed by Combustion Equipment Associates, Inc./Arthur D. Little, Inc. for Southern Services, Inc. Initial startup of the system occurred in February 1975; after the shakedown period, the EPA test program commenced in mid-May, 1975.

This paper presents a description of the system and its performance during the first year of operation from the initial startup through early January 1976, when the boiler was shut down for a scheduled overhaul. The boiler and the control system are due to be put back in operation by early March, 1976.

Preceding page blank

ACKNOWLEDGEMENTS

The authors would like to acknowledge several individuals for their invaluable contributions to the test program for the prototype dual alkali system.

We greatly appreciate the efforts of Indrakumar Jashnani, Bernard Jackson, and James Valentine of ADL who helped coordinate and supervise the test program at Scholz. Steve Spellenberg and Bruce Goodwin of ADL provided the bulk of the on-site chemical analytical support.

The cooperation of Combustion Equipment Associates, Gulf Power Company, and Southern Services, Inc. in this prototype test program is also gratefully acknowledged. In particular, we would like to thank Richard White of Combustion Equipment Associates, Reed Edwards of Southern Services, and James Kelly of Gulf Power for their on-site assistance in the operation of the system. Their continuing concern and attention to the various aspects of the test program and the system operation have been valuable contributions to the success of our overall efforts.

OPERATING EXPERIENCE--CEA/ADL DUAL ALKALI PROTOTYPE SYSTEM AT GULF POWER/SOUTHERN SERVICES, INC.

I. SUMMARY

During the first year of operation after startup, the prototype dual alkali system operated at an SO_2 concentration below the design range and at oxygen concentrations which exceeded the design range until the last three to four months of operation. Over the course of the year, the process and equipment performance improved significantly with excellent system availability during the final operating period from mid-September 1975 through early January 1976. The process has demonstrated very good reliability, stability of operation, and resistance to upset with no incidents of scrubber or mist eliminator plugging or scaling, even during the last three months of operation with no mist eliminator wash.

Under these conditions, the system operation relative to important performance parameters was as follows:

- SO_2 Removal--The system generally operated at SO_2 removal efficiencies of about 95% and demonstrated the capability for removal of over 99%. The amount of SO_2 removed is controllable by adjustment of scrubber pH.
- Oxidation and Sulfate Formation and Control--By co-precipitation of sulfate with sulfite, this concentrated dual alkali mode can keep up with oxidation rates of up to 25-30% in closed-loop operation.
- Lime Utilization--Lime utilization typically ranged from 95% to 100%. The overall lime stoichiometry was 0.95-1.00 moles $Ca(OH)_2$/mole of SO_2 removed from the flue gas.
- Waste Solids Properties--The system produced a washed filter cake generally containing at least 50% insoluble solids in this low to medium sulfur coal/high oxidation situation. The waste material, on the average, contained 3-5% soluble solids (dry cake basis, Period 2) with soluble solids reduced to the 2-3% range when washing was well controlled. Under process conditions consistent with higher sulfur coal operation, the insoluble solids content of the cake improved markedly with improved washability and lower soluble solids content.
- Sodium Makeup Requirements--About 0.03 moles Na_2CO_3 are required per mole of SO_2 removed as makeup for soluble solids losses in the washed cake under these process conditions. About an equal amount of sodium makeup was unaccounted for, probably due to leakage from the system and errors in soda ash makeup estimates. An attempt will be made to reduce leakage by maintenance performed during the shutdown period in January 1976.

Generally, the system performance has exceeded expectations for these flue gas conditions.

It is planned to put the system in operation again by early March 1976 and continue operations through about June 1976. During this next operating period, high sulfur coal (3.5-4.5%) will be in use at Scholz, enabling evaluation of the system performance in a high sulfur coal application.

II. INTRODUCTION

The 20-megawatt, prototype dual alkali process was designed and installed by Combustion Equipment Associates, Inc. (CEA) and Arthur D. Little, Inc. (ADL) at the Scholz Steam Plant of Gulf Power Company near Sneads, Florida. This system is one of three "second generation" 20-megawatt, prototype flue gas desulfurization systems installed at the plant as part of a technology evaluation program being conducted by Southern Services, Inc. for The Southern Company (an electric utility holding company including Alabama Power Company, Georgia Power Company, Gulf Power Company, Mississippi Power Company, Southern Electric Generating Company and Southern Services, Inc.).

The process was developed and designed jointly by CEA/ADL. Early laboratory research on the process, performed by ADL and sponsored by the Illinois Institute for Environmental Quality, dealt exclusively with characterizing the nature of the regeneration reaction. Based upon the laboratory results, a 2,000 cfm* dual alkali pilot plant was conducted at ADL's facilities by CEA/ADL and an eight-month test program was conducted to generate the design data for the prototype system. The pilot system contains the complete dual alkali process loop involving: gas scrubbing; absorbent regeneration; and solids separation. Results of the laboratory program and pilot operations for generation of the prototype system design have been reported previously in the literature.[1]

The laboratory and pilot plant investigation of dual alkali technology has continued at ADL in a program for the U.S. Environmental Protection Agency's (EPA) Industrial Environmental Research Laboratory at Research Triangle Park, N. C. The program involves characterization of the basic process chemistry and the various modes of operation of sodium-based dual alkali processes. The work covers a wide range of flue gas conditions, liquid reactant concentrations, and process configurations including the use of both lime and limestone for regeneration of the sodium scrubbing liquor. The three tasks of this program are described briefly below. Interim results were presented at the 1974 EPA Flue Gas Desulfurization Symposium.[2]

In Task I, the ADL Laboratory Program, experiments were performed on the regeneration of concentrated sodium scrubbing solutions using lime regeneration, limestone regeneration and sulfuric acid treatment for sulfate removal. Work was performed at EPA's laboratories at Research Triangle Park on regeneration using limestone in dilute mode operation.

---

* Despite EPA policy, certain non-metric units are used in the paper; however a conversion table is provided at the end of the paper.

Results of this prior work have been reported.[2] More recent work still in progress includes a characterization of the chemical, physical, and crystallographic properties of dual alkali solids, including work on the nature of calcium sulfate species co-precipitated with calcium sulfite in dual alkali processing. Work on physical properties includes evaluation of the mechanical and engineering properties of untreated dual alkali solids and solids produced from a limited number of mixtures of dual alkali solids with fly ash, lime and other agents used in the treatment of solid wastes from conventional lime/limestone flue gas desulfurization processes. This work is scheduled to be completed and reported in mid-1976.

Task II, the Pilot Plant Program, was conducted at the CEA/ADL pilot facility, Cambridge, Massachusetts. Results of short-term pilot operations using concentrated sodium scrubbing solutions with lime regeneration, limestone regeneration and sulfuric acid treatment for sulfate precipitation were reported previously.[2] Since then, short-term operations have also been conducted using dilute sodium (Na^+ active) scrubbing solutions with lime regeneration; and additional pilot operations have been conducted in an attempt to develop a successful dual alkali system using limestone for regeneration.

From the short-term pilot operations, three of the most promising modes of dual alkali processes were selected for long-term (three to five weeks around the clock), closed-loop pilot runs. These long-term tests were conducted in the following dual alkali modes:

- Dilute mode using lime for regeneration
- Concentrated mode using lime for regeneration
- Concentrated mode using limestone for regeneration

Work on the dilute and concentrated modes using lime has been completed. The results indicate that these modes can be operated closed loop with the following general performance:

- SO_2 removal - 90% or greater
- Lime utilization - 95% or greater
- Waste cake solids content - 45% or greater
- Sodium makeup requirements - less than 0.05 moles Na_2CO_3 per mole of SO_2 removed

The actual performance of the dual alkali process will vary depending upon the SO_2 and oxygen concentrations in the flue gas, the design of the system and the concentration of sodium solutions used in the process. Some version of the dual alkali process can generally be designed to exceed many or all of the above performance characteristics in most utility applications.

Aside from concentration differences, the principal difference in operating characteristics between dilute and concentrated lime dual alkali systems is that the dilute systems operate at or near saturation in calcium sulfate, which requires the use of carbonate makeup to provide some softening of the regenerated solution prior to recycle back to the scrubber.

Operations using limestone for regeneration have not been completed. Recent results have been promising, with many of the performance characteristics approaching those of lime dual alkali systems. The problem which remains to be solved is the production of solids with good settling characteristics over a wide range of sulfate, magnesium and iron concentrations in the scrubbing liquor.

The results of these long-term runs and recent work on limestone regeneration will be presented in a report scheduled for mid-1976.

Task III, the Prototype Test Program, consists of a one-year test of the 20-megawatt CEA/ADL prototype dual alkali system at Gulf Power. The objective of the program is to characterize the important aspects of the prototype process performance:

- SO_2 removal efficiency
- Oxidation and sulfate formation and control
- Lime utilization
- Waste solids properties
- Sodium makeup requirements and degree of closed-loop operation
- Process reliability

The prototype system was completed and put in operation by CEA/ADL in early February 1975. In mid-May 1975, the test program was initiated by ADL, Southern Services and CEA as part of the EPA/ADL Dual Alkali Program. The system was operated until early January 1976 when the boiler was shut down for a scheduled overhaul. The boiler and the prototype system are due to be put back in operation by early March 1976.

This paper describes the performance of the system over this first year of operation from startup in February 1975 through early January 1976. The startup period, including the first three months of operation prior to the EPA program, has already been reported in detail in the literature[3]; for completeness in describing the first year of operation, information on the first three months of operation is also included.

III. CEA/ADL 20-MEGAWATT DUAL ALKALI SYSTEM

A. PROCESS CHEMISTRY

The CEA/ADL dual alkali SO_2 control process at Scholz Station is a sodium solution scrubbing system in which the absorbent solution is regenerated using lime with the provision for use of limestone. The use of limestone is contingent upon successful pilot plant tests of a limestone mode.

The absorption of SO_2 is accomplished using a solution of sodium sulfite, sodium hydroxide and possibly some sodium carbonate (makeup) producing a spent sodium sulfite/bisulfite liquor:

$$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O \qquad (1)$$

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2\uparrow \qquad (2)$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \qquad (3)$$

During absorption, and to a lesser extent through the remainder of the system, some sulfite is oxidized to sulfate:

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4, \qquad (4)$$

converting an "active" form of sodium to an "inactive" form. Oxidation in the scrubber is generally a function of the scrubber design, oxygen content of the flue gas and the scrubber operating temperature. At excess air levels normally encountered in utility power plant operations burning medium or high sulfur coal, the level of oxidation is expected to be on the order of 5-10% of the sulfur dioxide removed.

The scrubber solution is regenerated by reaction with lime (and/or limestone) which precipitates a mixture of calcium sulfite and calcium sulfate solids for disposal, as shown by the following overall reactions (shown for lime):

$$2NaHSO_3 + Ca(OH)_2 \rightarrow Na_2SO_3 + CaSO_3 \cdot 1/2\ H_2O\downarrow + 3/2\ H_2O \qquad (5)$$

$$Na_2SO_3 + Ca(OH)_2 + 1/2\ H_2O \rightarrow 2NaOH + CaSO_3 \cdot 1/2\ H_2O\downarrow \qquad (6)$$

$$Na_2SO_4 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_4\downarrow \qquad (7)$$

After regeneration, the solids are separated from the regenerated liquor and washed. The clear liquor, containing very low amounts of suspended and dissolved calcium, is returned to the scrubbers. Soluble calcium levels are generally less than 100 ppm in concentrated dual alkali processes. Any sodium value lost with the washed waste solids is replaced by the addition of sodium carbonate to the regenerated liquor; sodium hydroxide can be used since carbonate softening is not required. Since sodium sulfate is also reacted with lime in this system to regenerate sodium hydroxide, it should be possible to use sodium sulfate as the sodium make-up source when low to moderate oxidation levels are encountered.

The system is designed to operate in the concentrated active sodium mode (active Na^+ concentration greater than 0.15M). In this mode, sulfate removal cannot be accomplished by the precipitation of gypsum ($CaSO_4 \cdot 2H_2O$), since the high sulfite levels prevent the soluble calcium concentration from reaching that required to exceed the gypsum solubility product. However, calcium sulfate ($CaSO_4$) is precipitated along with calcium sulfite ($CaSO_3 \cdot 1/2\ H_2O$) in the regeneration reactor, resulting in a solid solution of the two salts. The amount of sulfate precipitated in this form is a function of the concentrations of species in solution and the reactor pH. Under normal operation, with sulfate levels up to 1.5M $SO_4^=$, the system can keep up with sulfite oxidation rates equivalent to 25-30% of the SO_2 absorbed without becoming saturated in calcium sulfate.

Additional details of dual alkali chemistry and terminology are given in recent publications.[2,4]

B. SYSTEM DESCRIPTION

The prototype system at the Scholz Steam Plant is installed on Boiler No. 1, a 40-megawatt (nominal) Babcock and Wilcox pulverized coal-fired boiler. The prototype is sized to handle 50% of the flue gas from the boiler. The boiler has been retrofitted with a sectionalized, high efficiency electrostatic precipitator capable of 99.5% particulate removal. Sections of the electrostatic precipitator can be de-energized to study the impact of particulate input on process operation.

The design basis for the prototype system is given in Table 1; a schematic flow diagram is given in Figure 1. The system, designed as a prototype, incorporates a high degree of flexibility aimed at generating design and operating information for a wide variety of applications throughout The Southern Company system. Although the basic mode of operation of the system is a dual alkali process with lime regeneration, the system was designed to accommodate limestone regeneration and a combination of regeneration with limestone and lime. The system was also designed to enable operation as a direct limestone or lime scrubbing system. As a consequence, the system contains equipment and piping in excess of that required to operate a dual alkali system on a boiler already equipped with adequate particulate control.

TABLE 1

DESIGN BASIS

Flue Gas Inlet	
Flow Rate (acfm)	75,000
Temperature (°F)	275
O_2 Concentration (% volume, dry)	6.5 (max.)
Particulate Loading (gr/scf, dry)	0.02 (precipitator energized)
SO_2 Concentration (ppm, dry)	1800-3800
Design Performance	
SO_2 Removal (% reduction)	90 (min.)
Maximum SO_2 Removal Rate (lb/hr)	1530
Particulate (gr/scf, dry)	0.02 (no increase in loading)
Power Consumption (% power output)	
with venturi, full spray absorber pump requirements	2.5-3.0
without venturi, with tray absorber pump requirements	1.0-1.5

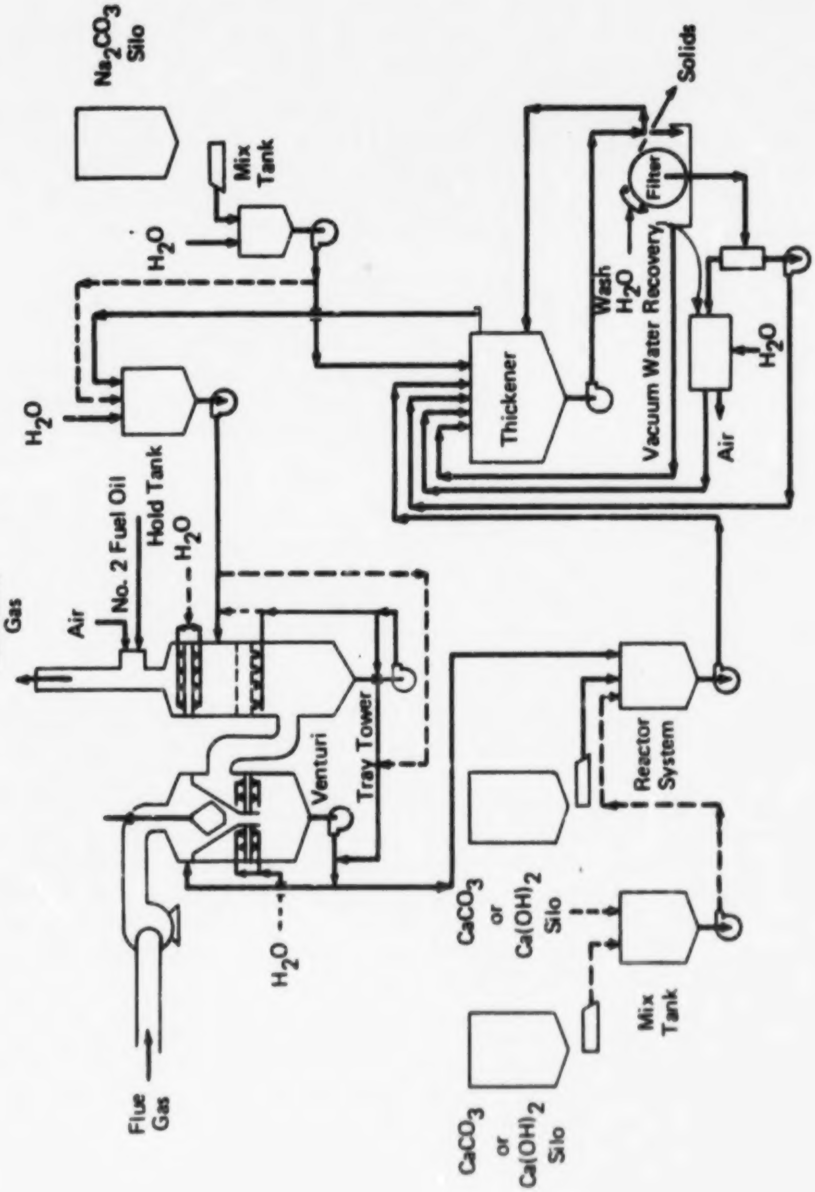

Flue Gas
H_2O
Venturi
Tray Tower
Scrubbed Gas
Air
No. 2 Fuel Oil
Hold Tank
H_2O
H_2O
H_2O
Mix Tank
Na_2CO_3 Silo
Thickener
Wash H_2O
Vacuum Water Recovery
Filter
Solids
Air
H_2O
Reactor System
$CaCO_3$ or $Ca(OH)_2$ Silo
Mix Tank
$CaCO_3$ or $Ca(OH)_2$ Silo

Figure 1. CEA/ADL dual alkali process flow diagram — Scholz Station.

The venturi scrubber is included in the system to investigate the operation of that type of scrubber in a dual alkali mode (or for direct lime or limestone scrubbing) when simultaneous particulate and SO_2 removal is required. The second scrubber, an absorption tower, can be operated as a tray scrubber, as in a dual alkali system where simultaneous particulate control is not required; or as a spray tower, for direct lime or limestone scrubbing, again without simultaneous particulate control. In addition to this redundancy in scrubbers, sufficient pump capacity is provided to operate the venturi at an L/G of 25 gallons/Macf of saturated gas and for an absorber L/G of 60, for operating in a spray tower configuration.

An additional storage silo (for limestone), a mix tank and other assorted tanks, pumps, controllers and piping were included in the system to accommodate the high degree of flexibility desired in the prototype system. To date, except when limestone has been erroneously delivered to the system instead of lime, the system has been operated only in the lime dual alkali mode as given in Figure 1.

In the lime dual alkali mode, the system design is based upon removal of at least 90% of the SO_2 in the flue gas for medium to high sulfur fuels (up to 5% sulfur) encountered in The Southern Company system. With the precipitator energized, the system was specified not to increase particulate loadings in the scrubber outlet above those in the inlet flue gas.

The power consumed by the system (not including reheat oil) is equivalent to 2.5-3.0% of the power generated by the unit in producing the flue gas load to the system, with the system operating at full fan capacity (full gas flow at a system pressure drop of 20 inches water) and at full venturi and absorber liquid recirculation capacity. Correcting for the excess fan and pump capacity, the power consumed by the equipment actually required in this application (tray tower at an L/G of 5-10) is roughly 1.0-1.5% of the power generated at the design conditions. In a full scale system designed for SO_2 removal only, the power consumption should be in this range.

The dual alkali system can be conveniently broken down into three process subsystems: gas scrubbing; absorbent regeneration; and solids dewatering. The design and operation of each of these subsystems is discussed in the following sections.

1. Flue Gas Scrubbing

The gas scrubbing system consists of a variable throat, plumb-bob type venturi scrubber followed in series by an absorption tower. Treated flue gas flows through both scrubbers. Each of these scrubbers is equipped with a removable liquid entrainment separator, an enclosed recycle tank to contain the scrubbing liquor, and two recycle pumps (one operating and one spare). The venturi can be used for absorption and/or particulate control and can be operated on a separate liquor loop from the absorption tower or in series with the absorber liquor loop. The absorption tower can be operated as a tray tower (with up to four trays), as a spray tower, or as a de-entrainment separator.

Gas is pulled from the exit of the electrostatic precipitator and forced through the scrubbing circuit using the booster fan provided with the dual alkali system. The fan and motor have been designed for a total system pressure drop of 20 inches H_2O at maximum flow. Under normal conditions with the precipitator in service, the venturi was used only for gas saturation with some SO_2 removal; a pressure drop of roughly 5 to 10 inches of water was maintained across the venturi.

Gas from the fan enters the venturi scrubber, flows downward over the wetted approach section and enters the high velocity venturi throat. Recycled scrubbing liquor also enters the top of the scrubber through tangential pipe inlets and through a number of vertical bull nozzles equally spaced around the center of the venturi.

After passing through the throat, the flue gas and scrubbing liquor continue downward through the internal downcomer; the liquor is collected at the bottom of the scrubber in the internal recycle tank. The flue gas makes a 180 degree upward turn and contacts the chevron-type entrainment separator (when used). The entrainment separator is equipped with wash sprays above and below which can be operated continually or on a sequential timing cycle. Either scrubbing liquor or fresh water can be used for washing. During dual alkali operations, this entrainment separator system was completely removed.

Following the venturi scrubber, the saturated flue gas enters the bottom of the absorption tower. This tower has been designed to operate either as a spray tower, with one or two sets of sprays, or as a tray tower, with from one to four trays. The bottom tray is equipped with a spray underneath to wet the bottom side of the tray. Gas passes upward through the trays and then through a final de-entrainment separator, with a wash system similar to the venturi demister wash. The wash system was used initially, but found to be unnecessary and its use discontinued.

The clean flue gas leaving the tower is finally reheated by the injection of hot gas from a burner fired with No. 2 fuel oil before being discharged through the stack on top of the absorber. Oil-fired reheat was specified by Southern Services for the prototype systems in order to reserve steam for power generation.

Regenerated absorbent solution, containing sodium hydroxide, sodium sulfite, sodium sulfate and some sodium carbonate, is fed to the top tray of the absorber. The solution flows countercurrent to the gas through the tray system (two trays) and is collected at the bottom of the absorber in the internal recycle tank. This collected liquor supplies solution both for spraying the bottom tray and for recirculation to the top tray, as needed for pH control in the scrubber or to maintain liquor flow across the trays. The pH in the combined stream of recirculated liquor and freshly regenerated feed ranged from about 6 to 11.

A bleed from the collected tray tower bottoms (pH roughly 5-7) is sent forward to the venturi recirculation loop for additional SO_2 removal. In the venturi, the gases with the highest SO_2 concentration are contacted with the most acidic liquor. A continuous bleed stream is drawn from the venturi recirculation loop and is sent to the absorbent regeneration system. The pH of the bleed stream normally ranged from 4.8 to under 6.0.

2. Absorbent Regeneration

Spent scrubber solution from the venturi recirculation line is bled to the regeneration reactor where it is reacted with hydrated lime. A full-scale system would normally incorporate a lime slaker and a slaked lime slurry tank; hydrated lime was used in this smaller installation for simplicity. The normal mode of operation was to feed dry lime directly to the reactor; however, provisions have been made to feed slurried hydrated lime as well as either slurried or dry limestone. The rate of the lime feed will in most cases be set according to the boiler load and coal sulfur content and adjusted on the pH of the reactor effluent liquor.

The lime neutralizes the bisulfite acidity in the scrubber bleed and further reacts with sodium sulfite and sodium sulfate to produce sodium hydroxide. These reactions precipitate mixed calcium sulfite and sulfate solids resulting in a slurry containing up to 5 wt % insoluble solids. The regeneration was normally carried out to a pH of between 11.0 and 12.5.

The regeneration reactor system consists of a multistage design, patented by CEA/ADL. The system at Scholz consists of two reactor stages in series: a short residence time reactor followed in series by a longer residence time reactor, designed for reactant residence times of 5 minutes and 35 minutes respectively at the design SO_2 removal rate (185 gpm flow rate through the reactor system). This reactor system design has been shown to generate clusters of sulfite/sulfate crystals which are generally spherical rather than the needle-like or platelet crystals generally associated with calcium sulfite precipitation. These crystal clusters have good settling, filtration and washing properties and can be generated in the reactor system over a wide range of flue gas and process conditions.

3. Solid/Liquid Separation and Solids Dewatering

Slurry from the regeneration reactor system is fed to the center well of the slurry thickener. The 40 foot diameter thickener is sized to handle the solids produced from the treatment of flue gases from the full 40 megawatts of Boiler No. 1.

The thickened slurry from the bottom of the settler is sent to a rotary drum vacuum filter. The solids content of the underflow was maintained below 30% to ease underflow pumping. The slurry is recirculated past the filter in a recycle loop that returns the slurry to the solids zone in the settler. The feed to the filter is drawn as the bleed from this recirculation loop. The filter surface area is 75 ft^2.

The solids cake is washed on the filter using two or three water sprays arranged in series. This wash removes a large fraction (up to 90%) of the occluded soluble salts from the cake and returns these salts to the system, thereby reducing sodium losses and minimizing sodium carbonate makeup.

Solids from the filter are retained in a solids storage enclosure under the filter from which they are loaded into dump trucks for transfer to the disposal area. The mixed filtrate and wash liquor from the filter are returned to the thickener.

Makeup sodium carbonate may be fed to the thickener in order to allow easy removal of any $CaCO_3$ precipitated. This carbonate is not intended for use as a softener, since soluble calcium concentrations in the regenerated liquor generally will run less that 100 ppm. However, some $CaCO_3$ precipitation will occur due to its very low solubility limits. The amount of $CaCO_3$ formed will be very small, a few hundred pounds per day at design conditions, which is less than 0.5% of the total cake produced.

Clear liquor overflow from the thickener is collected in the thickener hold tank which acts as surge capacity for the absorbent liquor feed to the scrubber system. Water can also be added to this hold tank to make up for the difference between total system water losses (evaporation and cake moisture) and total water inputs from other sources (sodium makeup solution, pump seals, lime feed, cake wash, and demister wash).

The disposal area for the dual alkali waste cake is a one-acre pit (450 ft x 100 ft) approximately 12 ft deep. The bottom and sides of the pit are lined with a layer of clay covered by a double thickness of polyethylene liner. The polyethylene is reinforced between the sheets with a mesh of nylon fibre. The floor of the pit slopes to a single collection drain constructed of PVC from which leachate is discharged to the ash pond. On top of the polyethylene liner is a two-inch layer of sand which gradually changes to gravel near the drain.

IV. OPERATING HISTORY

A. GENERAL OPERATING CONDITIONS

The prototype system commenced process startup on February 3, 1975 and was shut down on January 2, 1976 when the boiler was taken off the line for a scheduled overhaul. During this time the system was operated approximately 4,700 hours with shutdown periods of varying length for system maintenance and modifications and during periods when the boiler was taken off the line. Throughout the eleven months, the system treated flue gas with lower sulfur dioxide concentrations and higher oxygen concentrations than the range for which the system was designed. This represented a difficult test for the prototype system.

In sodium-based dual alkali systems for a given scrubber design and soluble solids level, the rate of sulfite oxidation (moles per unit time) is a strong function of the oxygen concentration in the flue gas and is relatively independent of the rate of sulfur dioxide removal (moles per unit time). As oxygen concentrations increase and SO_2 concentrations decrease, a higher percentage of the SO_2 removed from the flue gas is converted to sodium sulfate rather than sodium sulfite/bisulfite in the scrubber liquor. This higher percentage oxidation requires an increase in precipitation of sulfate relative to sulfite and a higher calcium sulfate content in the precipitated calcium sulfite/sulfate to enable closed-loop operation with no intentional purging of sodium sulfate.

Over the eleven-month period, sulfur dioxide concentrations in the flue gas averaged between 1,100 and 1,200 ppm, compared with a minimum design concentration of 1,800 ppm. The actual concentrations varied over a range of from 600 to 1,600 ppm, often fluctuating daily and even hourly over the entire range of concentrations. Oxygen concentrations in the flue gas entering the scrubber ranged from 5.0% to 10.5% (by volume equivalent to 30-100% excess air) with oxygen levels increasing with decreasing boiler load (higher excess air). During the first half of the year, air leaks in the boiler combustion air preheater and coal feed tubes contributed from 0.5 to 2.0 volume % to the oxygen concentrations. Oxygen levels in the flue gas were reduced from the 7-10% range down to the 5-7% range in September after air preheater repairs and when burner box pressure was brought under better control.

This general change in flue gas oxygen concentration and changes in the system operation made at about the same time logically led to the division of this first year of operation into two discrete operating periods described in the next section.

B. DESCRIPTION OF OPERATING PERIODS

1. Operating Period 1--Startup and Initial Operations

This first operating period extends from February 3, 1975 through July 18, 1975, when the system was shut down for a two-month period for modifications, repairs and to await receipt of replacement parts. During this operating period, commencing on initial process startup, the system was operated approximately 2,600 hours, 83% of the time that the boiler was in operation. Almost the entire system down-time of 518 hours relative to the boiler is accounted for by a three-week shutdown (491 hours) in mid-April, after the start-up period, for necessary maintenance and equipment modifications and adjustments prior to the start of the formal EPA test program in mid-May. Equipment changes are described in Section VB.

Table 2 contains a summary of the fuel and flue gas characteristics encountered during this operating period. The boiler was fired with a combination of several medium and low sulfur coals ranging in sulfur content from 0.9 to 2.2 wt %. The weighted average sulfur content of the coal as burned was 1.6 wt %, producing an average SO_2 level in the flue gas of about 1,050 ppm. Since it was not possible to segregate and selectively fire the different coals, the SO_2 levels in the flue gas fluctuated daily and often hourly from a low of 600 ppm to a high of 1,550 ppm.

The flue gas load to the prototype system was kept within a range equivalent to 15-22 Mw. However, the boiler load varied from 15 megawatts to 40 megawatts with an attendant variation in excess air rates.

In addition to the unfavorable process conditions resulting from low sulfur coal and the high oxygen concentrations, there were frequent process upsets. These upsets included occasional carry-over of fly ash with the flue gas and the inadvertent contamination of the lime supply with limestone (caused by a mix-up in the lime and limestone deliveries to the prototype systems at the plant). While the presence of fly ash in the system caused little or no effect on operation or performance, the limestone did produce temporary changes in the process chemistry--particularly when a significant fraction of the calcium feed to the reactor was limestone.

The limestone usually appeared mixed with lime at levels of up to 50% of the total feed; however, on a number of occasions (for periods lasting up to one day) the feed to the reactor system was pure limestone. During periods when only limestone was fed to the reactor, the pH in the reactor system dropped, SO_2 removal decreased somewhat and waste cake properties deteriorated slightly. Normal process conditions were re-established after the limestone passed through the system.

TABLE 2

SUMMARY OF SYSTEM INLET CONDITIONS

Operating Period 1 (2/75 - 7/75)

	Range	Average
Coal Fired:		
Sulfur (wt %)	0.9-2.2	1.6
HHV (Btu/lb coal)	11,500-13,000	12,000
Chloride (wt %)	0.05-0.10	0.08
Inlet Gas:		
Gas Load Treated (Mw equiv)	15-22	17
SO_2 Level (ppm-dry basis)	600-1550	1050
O_2 Level (% dry vol.)	5.0-10.5	7.5

There was a general rise in the flue gas oxygen concentration over the operating period due to worsening leakage in the boiler air preheater section. By mid-July oxygen concentrations were in the 8-10% range, with dilution of the already low SO_2 levels down to about 850-950 ppm. At the resultant high oxidation rates (30-50% of SO_2 removal), active sodium levels were inadvertently allowed to drop below 0.15M, into the range of dilute mode dual alkali operation. Under these conditions, soluble calcium levels in the regeneration system rose and the regenerated liquor became saturated in calcium sulfate, resulting in some gypsum scaling of the piping at the outlet of the regeneration reactor. At this same time, mechanical problems in the scrubber system (failure of two control valves and a pin hole leak in the absorber recycle tank) required that the system be shut down for repairs. After shutdown it was decided to await the repair of the preheater leak rather than attempt to continue operating at the low SO_2 concentrations and increasing oxygen concentrations.

Based upon this experience, it was decided that in future operations the active sodium concentration would be maintained well above the 0.15M level by operating with a slightly higher total sodium concentration in the system. This would allow the sodium sulfate concentration to increase in the system during periods of increasing rates of oxidation. Sulfate levels would continue to rise relative to the sulfite until the rate of sulfate precipitation as a calcium salt equaled the rate of oxidation. This can be accomplished without reverting to dilute mode operation at oxidation rates up to about 25-30% of the SO_2 removal rate.

The system was shut down from mid-July until mid-September. Repairs and revisions made during this period were of a mechanical nature rather than involving process changes and are discussed later. About two thirds of the shutdown period was to await replacement parts for the valves that had failed.

2. Operating Period 2--Low to Medium Sulfur Coal Operation

The system was put back in operation on September 16, 1975. From mid-September to mid-October repairs were made to the air preheater during boiler outages and adjustments were made in the boiler operation, reducing flue gas oxygen levels down to the 5-6% range. For the remainder of the test period through January 2, 1976, oxygen concentrations were generally kept in the 5-6% range. As shown in Table 3, SO_2 levels were slightly higher than those encountered during the first operating period, with an average level of about 1,200 ppm. Similarly, the gas load to the system was in the same range as that for Period 1. Active sodium concentrations were maintained in a range for concentrated mode operation and continuing improvements were made in the mechanical performance of system components, particularly the filter. One unprogrammed period of regeneration with limestone occurred in this interval.

TABLE 3

SUMMARY OF SYSTEM INLET CONDITIONS

Operating Period 2 (9/75-12/75)

	Range	Average
Coal Fired:		
Sulfur (wt %)	1.5-3.1	2.1
HHV (Btu/lb coal)	11,900-14,100	13,000
Chloride (wt %)	0.02-0.14	0.08
Inlet Gas:		
Gas Load Treated (Mw equiv)	16-19	18
SO_2 Level (ppm-dry basis)	800-1700	1220
O_2 Level (% dry vol.)	4.5-9.5	6.0

During this period, until shutdown on January 2, 1976, the prototype system operated approximately 2,100 hours, 97% of the time the boiler was in operation.

C. AVAILABILITY--FIRST YEAR PROTOTYPE SYSTEM OPERATION

The objective of the installation of the prototype system was to test the process chemistry and design on this relatively small scale in order to evaluate the viability of the process technology. While the reliability of the process was a principal concern, the 20-megawatt system was not intended to be a demonstration unit nor a test of the ultimate availability of such systems when applied full scale.

The system design was based upon scale-up by a factor of about 40 from the CEA/ADL dual alkali pilot plant. Although the prototype system contains spare pumps in the scrubber and regeneration areas, other key elements of the system such as the reactor system, lime feed system and filter station were not spared in this prototype application. Multiple installations and/or spare capacity for these critical items would normally be incorporated in full-scale applications. In spite of these considerations and the difficult flue gas conditions encountered during the first operating period, the availability of this prototype unit from initial startup through the first year of operation is impressive.

Boiler and prototype system operating hours are graphically displayed by month in Figure 2 and summarized by operating period in Table 4. The availability during the first operating period, 83%, includes the initial startup of the system and the three-week shutdown period for system adjustments and modifications prior to the start of the EPA test program. Such adjustments would normally be expected after process startup. During the second operating period, system availability exceeded 97%.

During the two months between operating periods, the system was down for maintenance and awaiting receipt of spare parts and repair of air leakage at the boiler air preheater. Including this period in the availability calculation for this first year yields an overall system availability of 69.3%. Exclusion of this interim period from the calculation yields availability during operating periods of 89%.

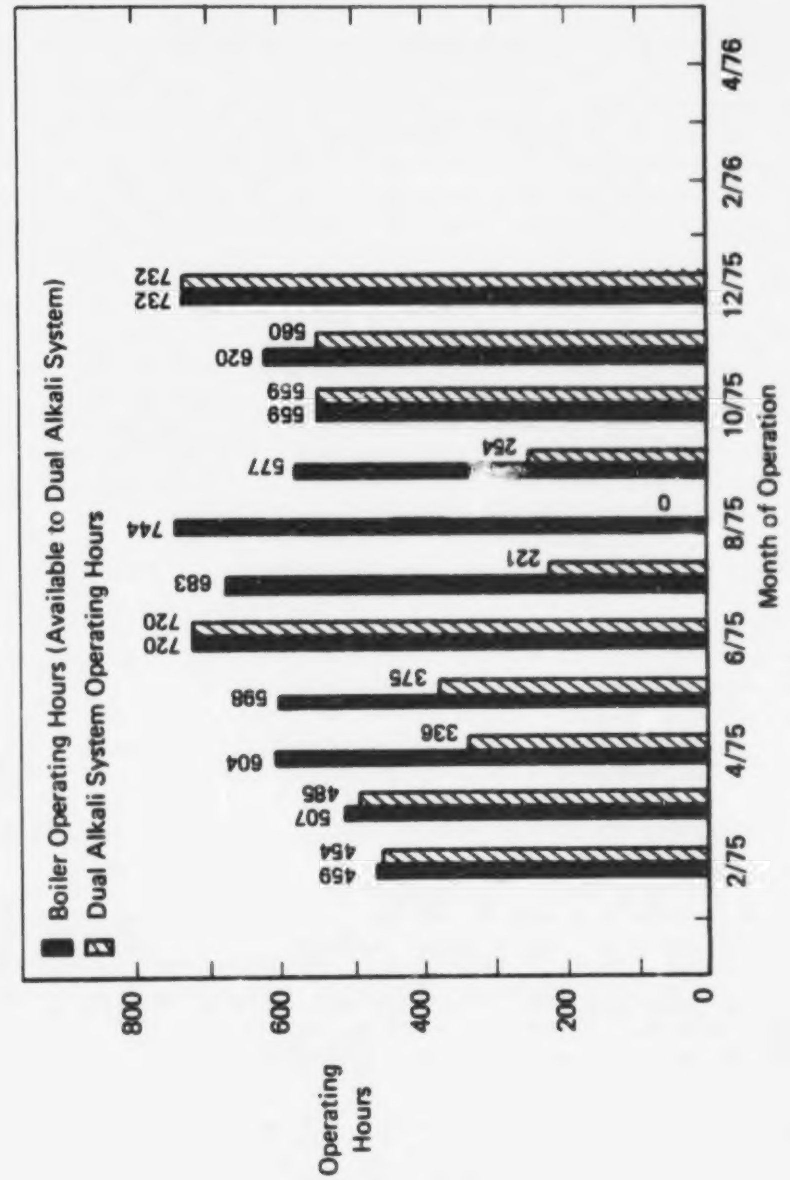


Figure 2. Availability of the CEA/ADL dual alkali system at Scholz.

TABLE 4

PROTOTYPE AVAILABILITY SUMMARY

	Dates	Operating Hours		Availability
		Prototype	Boiler[a]	Prototype Hours as percent of Boiler Hours
Operating Period 1	2/3/75-7/18/75	2591	3109	83.3%
Interim Period	7/18/75-9/15/75	0	1529	0
Operating Period 2	9/16/75-1/2/76	2153	2213	97.2
TOTAL YEAR		4744	6851	69.2
TOTAL OPERATING PERIODS		4744	5322	89.1

[a] Hours boiler available to dual alkali system.

V. SYSTEM PERFORMANCE

A. EQUIPMENT PERFORMANCE

The equipment performance in terms of overall reliability is reflected in the system availability and the level of maintenance required. Even though the availability of the system has been very good, there were numerous equipment and instrumentation problems. To a large degree, these problems were mechanical in nature and generally reflected equipment design or fabrication oversights commonly associated with a first-of-a-kind prototype system. There were only a few problems encountered that reflected process chemistry, and these required simple operational and/or equipment adjustments.

The appendix contains a detailed list of mechanical equipment and instrumentation problems encountered since startup. Included in the list are a number of problems of the type that would normally be expected to occur during startup or items normally requiring maintenance. The fact that some of the problems shown were not immediately corrected is an indication that they were either of minor significance or did not cause important operational problems. The more important problems are discussed below.

1. Equipment Problems

The mechanical equipment problems and maintenance items of primary concern are discussed below.

Filter. The filter was the largest source of problems in the prototype system. Because of anticipated corrosion problems associated with high chloride levels, a large part of the filtration equipment was fabricated out of fiberglass, which is not as sturdy as stainless steel and more prone to failures at stress points in the construction. Filter problems during Period 1 included:

- erosion of the fiberglass scraper blade, resulting in jagged edges which tore the cloth
- erosion of the bridge valve due to solids carried through cloth holes
- loss of vacuum due to cracks in the internal drum trunnion tubes
- cracking of the plastic caulking strips, allowing retention ropes to loosen and releasing the cloth panels
- failure of the fiberglass rocker arm used to agitate the slurry in the filter tub

Modifications were made to the filter during the latter part of Period 1 and the early part of Period 2. These modifications included replacement of the scraper blade with one fabricated out of stainless steel, reinforcement of the rocker arm with stainless steel plates, and the design of a new method for retaining the filter cloth panels. These changes along with regular inspection and maintenance and increased familiarity of Gulf Power personnel with the equipment have significantly improved filter performance.

Reactor System. Two mechanical equipment problems were encountered in the reactor system operation: plugging of the dry lime feed chute and solids buildup in the first reactor. The chute plugging problem was caused by hot vapors drifting up the chute entrance port and wetting the lime. The problem was resolved by installation of a vibrator on the chute with provision for injecting air to prevent vapors from rising into the chute.

The solids buildup in the first reactor was related to the use of dry lime feed (as opposed to slurry lime), feed chute location, and the poor agitation in the first reactor. These caused a deposition of solids both above and below the liquid surface, particularly in the area near the feed chute. A simulation of the conditions in this reactor at the CEA/ADL pilot plant in Cambridge has been successful in confirming the source of the solids buildup. A new first reactor has been designed and fabricated, and will be installed prior to restarting of the system. The lime slurry feed system is also being activated as an alternative to the feeding of dry lime. Feeding a slurry of slaked lime would be the normal practice in a full-scale application.

Scrubber. The rubber linings in the bleed control valves on both the absorber and venturi failed during operations in Periods 1 and 2. These failures may have been due to throttling to control flow. These valves were originally sized to accommodate direct slurry scrubbing at much higher flow rates. These rubber lined valves are scheduled to be replaced with stainless steel before the system is put back into operation.

Leakage of solution through the venturi and absorber pump seals has been of concern primarily because it represents a loss of sodium from the system. Tests were conducted by Southern Services and Gulf Power personnel to determine the size of the leaks under various pump packing conditions and to determine how best to eliminate or minimize the leaks. A Teflon impregnated packing was installed just prior to shutdown and it appears to have all but eliminated the leaks. However, further testing with this packing is required.

A slight corrosion and ash buildup on the fan during extended shutdowns was diagnosed to be due to flue gas leaks through the isolation damper. Since the fan is located upstream of the system it is constructed out of carbon steel and significant corrosion was not expected. The corrosion and ash buildup was generally minor and the fan rotor was easily cleaned and rebalanced prior to startup. Correcting the leak through the isolation damper by installing a new damper or use of an air seal system was considered to be unwarranted.

Thickener. The major problems with the operation of the thickener were plugging of the underflow lines (Period 1) and leaking of liquor through the bottom due to lining failure (Period 2). Plugging of the thickener underflow lines was partly due to the design of the underflow piping, and partly due to the frequent downtime on the filter which allowed thickener underflow slurry concentrations to exceed 30% solids. Redesign of the underflow lines using flexible piping and adjustments to the operational procedures to maintain the underflow slurry concentration in the 15% to 25% range (by back-flushing with clear hold tank liquor to dilute the underflow when necessary), have effectively eliminated underflow plugging problems.

The leak that developed in the bottom of the thickener during the early part of the second period of operation was small but grew worse throughout Period 2. After shutdown in January all the solids were pulled from the thickener and the liquor drained. Inspection of the inside of the thickener showed sections of lining to be failing due to poor curing or poor application by the lining supplier. The leaks have been located and patched, and the sections of failed lining are being replaced.

Interestingly, when the thickener was drained, pieces of rope and a flattened paint can were found in the area of the cone. These items undoubtedly contributed to difficulties with the thickener underflow system. A piece of rope had previously been extracted from a seized underflow pump during operation in Period 2.

2. Instrumentation

Instrumentation problems primarily involved the pH units, level transmitters, and soda ash feed solution control system. Other instrumentation problems occurred, but for the most part, these were minor.

Flow-through pH units were originally installed throughout the system. The piping for these units had a tendency to plug and the electrodes coat with a fine film, causing a drift in pH readings. In the case of the pH unit in the reactor system, the lines and probe chamber plugged completely with solids at low slurry flow rates and the probes eroded away at high flow rates. The flow-through unit on the reactor system was replaced with an immersion-type unit fitted with a sonic cleaner midway through Period 1. This has proved to be much more reliable; however, there have been problems keeping the sonic cleaner mounted on the probe casing. The unit now requires routine checking and recalibration about every week to two weeks.

The take-off lines for the flow-through units on the scrubber system have been or are now being relocated. Also, somewhat higher flow rates will be maintained through the probes to prevent plugging of piping.

Since close pH control is not required, the problems with the pH units have not been critical. Also, the scrubber system can be operated on either the venturi bleed pH or outlet SO_2 (or for that matter, inlet SO_2

and feed forward flow) with only occasional checks of the bleed pH to verify the SO_2 monitor. In fact, during a few weeks in the early part of Period 1, when all pH units and SO_2 monitors were out of service (due to delays in obtaining replacement parts), the system has been successfully operated by taking hourly pH readings with a portable pH unit.

The level transmitter/controllers on a few tanks have been unreliable. The particular type of unit could not be completely serviced in the field. Thus, there were periods when inaccurate level indication and/or control have caused operational problems. These units have been replaced.

Similarly, the soda ash solution feed control system has also been unreliable. This, in combination with occasional plugging of the feed gate on the dry soda ash feeder, has made it difficult to close the overall material balance on sodium. During the last half of the second operating period, the sodium makeup rate was determined using frequent checks on the specific gravity of the soda ash solution and recalibration of the flow indicator along with the inventory of soda ash silo and the quantity of soda ash delivery. Inventory and delivery information alone would not have been sufficient on the short-term due to the large storage capacity in the silo and the low soda ash feed rates. The soda ash feed control system is now being revised.

B. PROCESS PERFORMANCE

The general process conditions for the two periods of operation are summarized in Tables 5 and 6. During Period 1 (February 3--January 18, 1975) the inlet SO_2 concentration ranged from 600 to 1,550 ppm and averaged about 1,050 ppm. Flue gas oxygen levels varied from 5% to about 10% and typically fluctuated between 6.5% and 8.5%.

The system operating philosophy during Period 1 was to maintain the active sodium concentration in the system at or above 0.2M Na^+ and to maintain a total sodium concentration of about 2.0M. This resulted in a fairly stable operation until the end of June, when the inlet SO_2 level fell to 850-950 ppm and oxygen levels rose to 8.5-10.0%. Under these conditions, without an increase in the sodium carbonate feed rate, the system chemistry drifted into the range of a dilute dual alkali mode with active sodium concentrations dropping to below 0.15M Na^+. Except for the two-week period of dilute mode operation in July, the thickener liquor typically contained 0.20-0.25M active sodium, 0.7-0.9M sodium sulfate, and 0.10-0.15M sodium chloride (4,000-5,500 ppm Cl^-).

At the start of Period 2, it was decided to maintain the active sodium concentration in the system above 0.3M Na^+ and to allow the sodium sulfate concentration to fluctuate to any level necessary to maintain an equilibrium between sulfite oxidation in the system and sulfate precipitation. This would prevent any possibility of deterioration of the system chemistry to that of a dilute mode.

TABLE 5

SUMMARY OF SYSTEM OPERATING CONDITIONS

Period 1
(2/3 - 7/18/75)

	Range	Average
Inlet Gas:		
Gas Load Treated (Mw equiv.)	15-22	17
SO_2 Level (ppm - dry basis)	600-1,550	1,050
O_2 Level (% dry vol.)	5.0-10.5	7.5
	Range[a]	Typical[a]
Regenerated Liquor Composition:		
pH	10-12.6	11-12.5
Na^+_{act} (M)	0.10-0.35	0.2-0.25
$SO_4^=$ (M)	0.60-1.05	0.7-0.9
Cl^- (ppm)	3,000-7,000	4,000-5,500
Ca^{++} (ppm)	20-800+[b]	50-200

[a] For March - July 1975, not including periods when limestone fed to reactor.

[b] Ca^{++} levels above 250 ppm occurred during the period just prior to and during the dilute mode operation.

TABLE 6

SUMMARY OF SYSTEM OPERATING CONDITIONS

Period 2
(9/18 - 12/25/75)

	Range	Average
Inlet Gas:		
Gas Load Treated (Mw equiv.)	16-19	18
SO_2 Level (ppm - dry basis)	800-1,700	1,200
O_2 Level (% dry vol.)	4.5-9.5	6.0
	Range[a]	Typical[a]
Regenerated Liquor Composition:		
pH	10-12.8	11-12
Na^{+}_{act} (M)	0.25-0.6	0.3-0.35
$SO_4^{=}$ (M)	0.6-1.05	0.8-1.0
Cl^{-} (ppm)	1,000-2,100	1,300
Ca^{++} (ppm)	30-160	70

---

[a] For October - December 1975, not including periods when limestone fed to reactor.

The flue gas conditions during Period 2 improved over those experienced in Period 1, particularly in regard to the oxygen levels. Inlet SO_2 concentrations ranged from 800 to 1,700 ppm and averaged about 1,200 ppm. Flue gas oxygen levels generally were controlled to 5-6% with only infrequent excursions above 6.5%.

At these flue gas conditions, with the active sodium concentration maintained above 0.3M, the sodium sulfate concentration typically ran 0.8-1.0M, and the chloride concentration fluctuated in the 0.03-0.06M range ($\sim$1,300 ppm Cl^-). Routine analyses of coal samples performed for the plant from both Periods 1 and 2 show about the same chloride content, approximately 0.08 wt % on the average. However, the accuracy of these analyses is only about ±0.05 wt %.

The various aspects of the process performance during the two operating periods are discussed in the following sections. The discussion has been broken down into six areas:

- SO_2 removal
- Lime utilization
- Oxidation and sulfate control
- Waste cake properties
- Sodium makeup, and
- Process reliability

Complete material balances could be closed only during selected intervals in each operating period because of the frequent contamination of the lime supply with limestone, fluctuating flue gas and operating conditions, and occasional instrumentation problems. However, sufficient analytical and operational data exist throughout both periods (even when material balances were not complete) to allow thorough characterization of the system operation and performance on the low and medium sulfur coal fired during the first year.

1. SO_2 Removal

The scrubber system was operated using two different configurations for SO_2 removal: the venturi and absorber together in series (with two trays) and the venturi alone. In the latter configuration the trays were not removed from the absorber; rather, the regenerated liquor feed to the top tray was diverted either to the absorber recycle tank (from which it was transferred to the venturi) or bypassed directly to the venturi through a line installed in May 1975. (The bypass line was not in the original design, since the operation on low sulfur coal was not anticipated.) The recycle flow to the top tray and the flow to the spray on the underside of the bottom tray were both discontinued; however, the absorber pumps were maintained in operation to transfer liquor collected in the absorber tank to the venturi. In order not to overly back-pressure the absorber pumps, a recycle was maintained through an open spray header during intervals when there was no feed to the trays.

During Period 1, both operational configurations were used at different times. However, during Period 2, only the combined venturi and absorber configuration was used.

The SO_2 removal efficiency achieved with each of these configurations is shown in Figure 3 for intervals when the inlet SO_2 ranged from 1,050 to 1,250 ppm, the typical inlet level. Points shown on this plot represent data taken during the normal course of operations in both periods when inlet and outlet SO_2 levels and system pH's were simultaneously available and confirmed. In most cases, each data point represents at least a few hours of operation at the condition shown. Where such a continuum of data exists, outlet SO_2 levels and pH's have been averaged and rounded to the nearest 5 ppm and 0.05 pH units, respectively.

The data in Figure 3, which reflects the general operating experience at Scholz, confirm the high SO_2 removal capability of sodium-based scrubber systems operating in an equivalent concentration range of active sodium. Achieving a given outlet SO_2 level (within the limit of the number of contact stages in use) was essentially a matter of adjusting the operating pH of the scrubber system by changing the feed forward rate and/or pH of the regenerated liquor.

Using both the venturi and absorber (with two trays), outlet SO_2 levels below 50 ppm could be easily achieved at a venturi liquor pH above 5.2. This corresponds to better than 95% removal efficiency. Higher removal efficiencies were attained with increasingly high pH levels; however, there was little to be gained by operating at pH's much in excess of 6.0. Above a pH of 6.0, outlet SO_2 levels dropped to 20 ppm or less (>98% removal), taxing the accuracy and operating range of the outlet SO_2 monitor.

For the most part, when both the venturi and absorber were operated together, the pH of the venturi bleed liquor was maintained between 4.8 and 5.9. Consequently, outlet SO_2 levels generally ranged between 15 and 100 ppm, and typically ran 30 to 70 ppm. In Period 2, for example, during December and November the outlet SO_2 level ranged from 5 to 210 ppm and averaged 45-50 ppm. During these two months there were only five occasions when the outlet SO_2 exceeded 100 ppm. Three of these occurred while the system was being operated with limestone rather than lime fed to the regeneration reactor.

Operating with the venturi alone (liquor bypassing absorber trays) a pH feed of 6.4-6.5 was required to achieve 95% removal (less than 50 ppm outlet SO_2). However, better than 90% removal could still be quite easily achieved at pH's on the order of 6.0. When the venturi alone was used in Period 1, the bleed pH was generally maintained above 5.7 to keep outlet SO_2 levels at or below about 100 ppm.

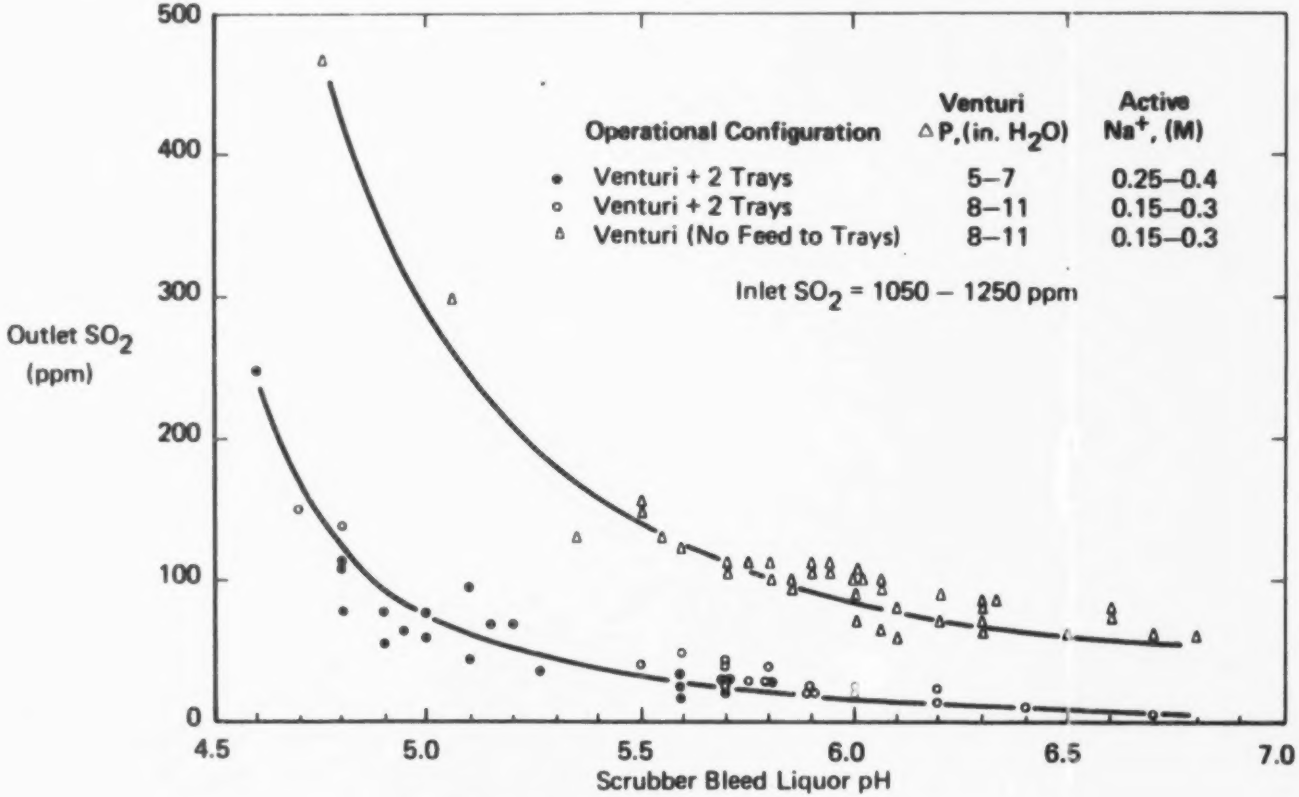


Figure 3, SO_2 Removal in the Scrubber System as a Function of pH of the Scrubber Bleed Liquor

2. Lime Utilization

In general, lime utilization was quite good, equalling that predicted from the laboratory and pilot plant developmental work. Under normal operating conditions, lime utilization ranged from 90% to 100% of the available $Ca(OH)_2$ in the hydrated lime (the hydrated lime ranged from 87% to 93% available $Ca(OH)_2$). Based upon analysis of the solids produced, lime utilization typically ran about 95% of the available $Ca(OH)_2$. These data are generally confirmed by overall system material balances.

During Period 1 it is not possible to accurately calculate the overall $Ca(OH)_2/\Delta SO_2$ stoichiometry due to the frequent operational upsets, including the occasional feeding of limestone. The best estimate for the first five months of operation (after initial startup) is 0.95 to 1.00. This estimate is based upon the integral averages of SO_2 inlet and outlet and lime feed (adjusted for estimated unreacted limestone). During November and December 1975, when accurate material balances were closed, the $Ca(OH)_2/\Delta SO_2$ stoichiometry was 0.98. Sodium carbonate makeup represented an additional alkali input to the system.

3. Oxidation and Sulfate Control

As would be expected, the major fraction of sulfite oxidation in the system occurred in the scrubber circuit; and the single most important variable influencing the oxidation rate was the oxygen concentration in the flue gas. Estimated oxidation rates in the scrubber circuit ranged from a low of about 150 ppm equivalent SO_2 (at an oxygen level of 5%) to a high of about 400 ppm (at an oxygen level of 9%). By contrast, the oxidation in the remainder of the process (the regeneration and dewatering systems combined) was less than 25 ppm. Thus, oxidation in the scrubber circuit accounted for more than 90% of the total system oxidation.

Figure 4 shows the estimated oxidation rates in the scrubber system during both periods in equivalent ppm of SO_2 (design gas flow basis) as a function of flue gas oxygen content. Included in this plot are data for both scrubber system operational configurations.

It would be expected that oxidation rates would be somewhat lower using the venturi alone than with the combined venturi and absorber, due to the decreased gas/liquid contacting. However, such a decrease in oxidation was not observed. This can be attributed to three factors: first, the carryover of entrained liquor from the venturi to the absorber; second, contacting of gas with liquor recirculated through the open spray header; and finally, the flow of a small amount of regenerated liquor onto the top tray through a leaky shutoff valve.

Considering all data, regardless of scrubber configuration, the data scatter exhibited in Figure 4 amounts to a range of 80 ppm of oxidation at a given flue gas oxygen level. This data scatter can be accounted for by differences in operating temperature (120-135°F), liquor flow, and slight differences in pressure drop in the tray tower, as well as sampling and analytical errors.

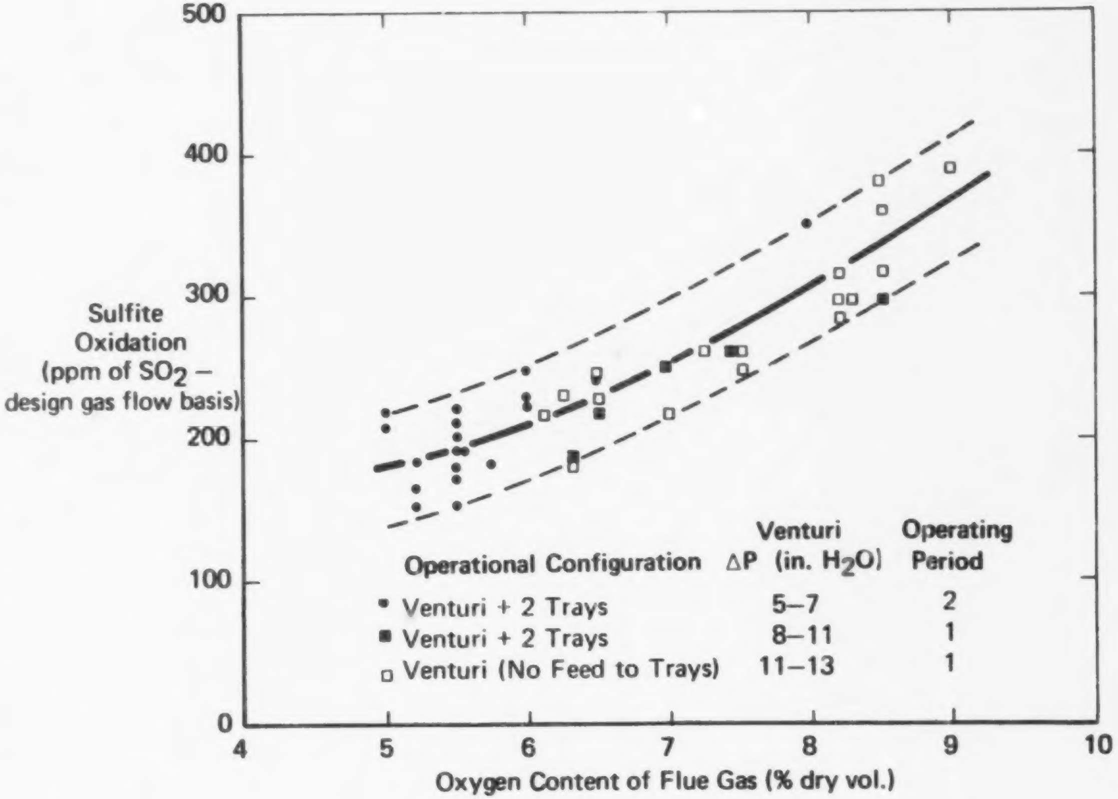


Figure 4, Oxidation in the Scrubber System as a Function of Flue Gas Oxygen Content

Assuming the median value of oxidation to represent the average oxidation experienced, then oxidation ranged from 175 ± 40 ppm at 5% oxygen to 380 ± 40 ppm at 9% oxygen. For a typical SO_2 level of 1,200 ppm with 95% SO_2 removal and 90% gas load, these oxidation rates correspond to 17% and 37% of the SO_2 removed. Adding in the oxidation through the remainder of the system, the total oxidation would be 20% of the SO_2 removed at a 5% oxygen level and 40% at 9% oxygen. For high sulfur coals (3-4% S), these same overall oxidation rates would correspond to roughly 8% and 16%, respectively.

It is worth noting that these oxidation rates represent operation with three active contact stages (apparently even when the venturi was operated alone). Normally in a low or medium sulfur coal application only one or two stages should be required. This reduction in the number of stages should substantially reduce oxidation rates.

At steady state the total sulfate formed must be removed from the system by the precipitation of calcium sulfate and/or sodium sulfate losses at the rate it is being formed. Calcium sulfate levels measured in the product solids ranged from as low as a few percent of the total insoluble calcium-sulfur salts during early startup to as high as 30% during periods of high oxygen levels and low SO_2 (when the ratio of sodium sulfate to active sodium in the system liquor was fairly high). The typical range of calcium sulfate in the waste cake during Period 1 was 15% to 25% (mole basis) of the total calcium-sulfur salts; and in Period 2 was 12% to 20% (mole basis). These levels of calcium sulfate precipitation indicate that the system is capable of keeping up with oxidation rates of up to 25-30% of the SO_2 removal in the concentrated mode of operation. In a situation where the cake was thoroughly washed, the bulk of the sulfate losses would be as calcium sulfate; sodium sulfate losses would represent less than 2-3% of the SO_2 absorbed. The system liquor composition would simply self-adjust the ratio of sodium sulfate to active sodium to precipitate the amount of calcium sulfate required to keep up with oxidation.

The sulfate balances established during November and December 1975 are shown in Table 7. These months represent stable operating intervals in which material balances were closed. During November, 20% of the SO_2 absorbed was oxidized to sodium sulfate in the system. Of this 20%, 13% left the system as insoluble calcium sulfate. The remaining 7% left as sodium sulfate. About half of the sodium sulfate was lost in the liquor occluded with the cake (∿5% sodium salts on a dry cake basis); and the other half through uncollected pump seal leakage, the thickener leak, and other unaccounted losses. The total sodium sulfate loss was calculated from the soda ash makeup during this period.

In December, 23% of the SO_2 absorbed was oxidized to sulfate. About 15% left as insoluble calcium sulfate; about 3% as sodium sulfate occluded with the cake (∿3.5% sodium salts on a dry cake basis); and the rest through system leaks and increase in liquor sulfate concentration.

TABLE 7

SULFATE BALANCES DURING PERIOD 2

	11/3-11/23/75	12/2-12/23/75
Average Inlet SO_2 (ppm)	1,265	1,135
Average SO_2 Removal (%)	95.5	96.5
Average Inlet O_2 (vol %, dry)	5.5	6.0
Total Sulfate Formation (% ΔSO_2)	20.0	23.0
Sulfate Account (% ΔSO_2):		
In Cake - $CaSO_4$	13.0	15.0
- Na_2SO_4	4.5	3.0
Na_2SO_4 in System Liquor Inventory	-0.5	2.5
Other[a]	4.0	3.0
Total	21.0	23.5

---

[a] This Na_2SO_4 loss is calculated from the difference between total net sodium makeup and sodium losses in the cake and represents sodium sulfate losses in pump seal leaks, spills, and errors in estimates of soda ash makeup.

Had all system leaks been stopped or returned to the system and the cake been washed to the same level of soluble solids, then the sulfate to active sodium ratio in the liquor would have increased somewhat to allow more calcium sulfate to precipitate, thereby re-establishing the equilibrium between sulfate formation and losses.

4. Waste Cake Properties

The solids content of the filter cake produced during both periods of operation typically ranged from 45% to 60% solids. In general, the solids content decreased with increasing calcium sulfate content of the cake. Process upsets (e.g., high levels of limestone in the lime feed) and mechanical problems with the filter also tended to depress the level of solids in the cake to the 45% range.

During intervals of normal process conditions with the filter operating under standard operating conditions, the solids content of the cake usually varied between 50% and 55% for calcium sulfate levels in the cake of between 10% and 25% of the total calcium-sulfur salts. At this level of solids, the cake generally had the appearance and consistency of a moist powder.

During startup, when active sodium levels were fairly high (above 0.4M) and sulfate concentrations were below 0.6M, the solids content of the cake ranged from 55% to 70%, indicating the potential for producing even drier cake under high sulfur coal conditions.

Wash efficiency tests conducted at Scholz during Period 1 indicated that with two banks of wash sprays in series, the soluble solids content of the cake could be reduced to 2-3% (dry cake basis) under controlled filter conditions using a wash ratio (gals. wash water/gal. water occluded in cake) of about 2.5. However, mechanical problems and process upsets frequently prevented a reasonably continuous operation of the filter with a thin enough cake to allow consistently high wash ratios and, therefore, low solubles losses. The levels of solubles in the filter cake in Period 1, therefore, ranged from 2% to 12% of the dry cake weight with an average of 5-8%. The 12% solubles level occurred during periods when the cake was not washed.

Prior to the start of the second period of operation and during the early part of Period 2, a third bank of wash sprays was added and the capacity of the existing sprays increased. This resulted in a doubling of the wash capacity from 3 to about 7 gpm, which ensured that sufficient wash water would be available during periods of high rates of cake withdrawal. In November, the level of soluble solids in the cake averaged 4.8% (dry cake basis) with a wash ratio of 1.5-2.0. In December, the wash rate was increased to a wash ratio of 2.0-2.5 and the average soluble solids level was decreased to 3.3% (a range of 1.5% to 6.0% dry cake basis). The average solids content during both months was 51%.

During Period 2, a limited amount of testing was performed with different types of filter cloths. A number of different nylon and polypropylene cloths in both monofilament and multifilament weaves were tried. The original cloth, a polypropylene multifilament, was durable. However, it produced poorer solids (45-50%) under the more adverse process conditions and tended to blind, requiring cleaning of the cloth about two to three times a week. The purpose of the testing was to eliminate or minimize the blinding while maintaining the high solids content and durability. To date, the best cloth has been a polypropylene monofilament with the same porosity as the original multifilament. This cloth was used exclusively in December 1975. The durability of the cloth is still being tested.

5. Sodium Makeup

The sodium makeup requirement in the operation of a dual alkali system is determined simply by the rate of sodium loss, both controlled (solubles in the filter cake) and uncontrolled (pump seal leaks, tank spills, etc.). Under normal operating conditions the losses in the filter cake should be the single most important sodium loss, and sodium makeup should equal or only slightly exceed the quantity contained in the cake (on a mole equivalent basis) in a tight, closed-loop operation.

Wash efficiency tests performed at Scholz during Period 1 indicated that 2-3% solubles losses in the cake could be achieved with a wash ratio of 2.5. This solubles loss translates directly into an equivalent soda ash makeup requirement of 2-3% of the SO_2 absorbed on a mole basis, which represents a reasonable lower limit on soda ash makeup.

Soda ash makeup to the system has been consistently higher than this level throughout the first year of operation. During the first half of the year (Period 1) mechanical problems with the filter and insufficient washing limited the level to which soluble sodium salts could be washed from the cake over an extended period to a minimum of about 5-8%. In addition to the cake losses, there were also inadvertent spills as new operators gained familiarity with the system as well as the normal leaks from pump packing, piping, etc. Thus, soda ash makeup requirements frequently exceeded 10% of the SO_2 absorbed on a mole basis.

The improvements in the filter operation and increased wash efficiency in Period 2 substantially reduced sodium losses in the cake to 4.8% solubles in November and 3.3% solubles in December. These sodium losses corresponded to a soda ash makeup requirement equivalent to 5.8% and 3.6% of the SO_2 absorbed (moles Na_2CO_3 per 100 moles of SO_2 removed), respectively. The net soda ash makeup (not including that used in increasing the inventory of sodium in the system), though, was 10.5% in November and 7% in December. The higher makeup requirement in November is partially due to contamination of the lime with 25 tons of limestone, increasing the amount of waste solids and associated sodium losses.

The apparent difference of 3-4% between the estimated soda ash feed and sodium losses in the cake may be due to leakage from the system (pump seals and the small thickener leak) and errors in estimates of the soda ash feed.

Entrainment losses of sodium were very small. In particulate sampling conducted during December, the total weight of sodium salts in the scrubbed gas averaged 0.002 grains/scf. This represents a liquor loss through entrainment of less than one gallon per hour. The soda ash makeup required to replace this entrainment loss is less than 0.1 mole % of the SO_2 absorbed.

Efforts will continue throughout the remainder of the testing to minimize soda ash makeup requirements by adequate washing of the filter cake and further control of the leakage of liquor. The better cake produced with high sulfur coal should allow reduction of solubles losses in the filter cake to below 3%.

6. Process Reliability

Process reliability refers to the overall ease of process operation including the resistance of the process chemistry to operational upsets, the sensitivity of the process performance to small changes in the process chemistry, and the potential for scaling in the process equipment. In this regard it is to be differentiated from the mechanical/equipment-related problems previously discussed.

In all respects the process reliability has been excellent. The system has been successfully operated over a range of widely fluctuating inlet SO_2 levels and oxygen concentrations in the flue gas, with little or no change in the SO_2 removal efficiency, cake properties or lime utilization. Throughout Period 1 with active sodium concentrations between 0.15 and 0.30M, soluble calcium levels generally fluctuated between 20 and 250 ppm and were usually below 150 ppm. In Period 2, when active sodium levels were maintained above 0.3M, soluble calcium concentrations averaged about 70 ppm and were consistently below 100 ppm. In this range of calcium concentration, 20-250 ppm, there were no problems with scale formation in the reactor (other than the mechanical problems already discussed), the dewatering, or scrubber systems.

The low potential for scale formation and solids deposition in the scrubber system was further demonstrated in Period 2 when the wash sprays beneath the demister were turned off. For the three-month period from October through December the demister was operated without wash water. There was no increase in pressure drop across the demister nor the formation of a scale or deposit of any kind.

The only period when difficulties were experienced due to changes in the process conditions was in July when the inlet SO_2 levels dropped to 850-950 ppm and the flue gas oxygen level rose to 8.5-10.0%. The system was allowed to drift into a dilute mode (active sodium levels decreased to less than

0.15M). The very low soluble sulfite concentrations caused the soluble calcium concentrations to rise to gypsum saturation levels and above, resulting in the precipitation of gypsum and the deposition of gypsum scale in the reactor vessels and reactor outlet piping. The scale was removed during system shutdown. Had the active sodium been increased by charging additional soda ash to the system, the gypsum scale would have slowly redissolved, as evidenced by the softening and eventual redissolution of gypsum scale that was not removed during the shutdown.

In addition to the fluctuating SO_2 levels and oxygen concentrations, the system also was operated during a number of process upsets. These upsets included the carryover of small amounts of fly ash (Period 1), the contamination of the lime feed supply with limestone (Periods 1 and 2), and overfeeding of lime (Period 1). Only the limestone had any effect on the process performance and this effect was temporary. During periods when pure limestone was fed to the reactor system, the pH of the regenerated liquor fell to below 7, causing a loss of SO_2 removal efficiency, from above 90% down to 80-85%. The solids content of the filter cake also decreased slightly and sodium losses in the cake correspondingly increased. Within a day or two after the limestone had passed through the system, all of the effects were reversed and the system operation was back to normal.

APPENDIX

EQUIPMENT/INSTRUMENTATION STARTUP AND MAINTENANCE PROBLEMS

System	Problem	Action	Status
SCRUBBER	EQUIPMENT:		
	• Deterioration of refractory in reheat burner chamber	Replaced	Corrected (Period 1)
	• Slight corrosion and ash buildup on fan during extended shutdown periods	Clean and rebalance fan after extended shutdowns	--
	• Separation of bond on rubber linings in absorber and venturi bleed control valves	Replace with SS valves under warranty	Scheduled (1-2/76)
	• Hairline cracks and pinholes in lining on absorber recycle tank	Patched	Corrected (Period 1)
	• Leaking of liquor through pump seals and piping on pump suction lines	Replaced packing and flange gaskets	Testing of new packing (Period 2)
	• Deterioration of fan thrust bearing	Replace	Scheduled (1-2/76)
	• Deterioration of stack lining due to poor curing	Patch/replace under warranty	Scheduled (1-2/76)
	INSTRUMENTATION:		
	• Water condensation in pressure taps on level controllers	Relocated lines	Corrected (Period 2)
	• Plugging of flow-through pH units on recycle tanks	Relocate sampling lines	Scheduled (1-2/76)
REACTOR	EQUIPMENT:		
	• Plugging of dry lime feed chute to first reactor	Installed vibrator	Corrected (Period 1)
	• Buildup of solids in first reactor due to location of dry lime feed chute and poor agitation	Design and install new first reactor	Scheduled (1-2/76)

EQUIPMENT/INSTRUMENTATION STARTUP AND MAINTENANCE PROBLEMS

umber	Problem	Action	Status
TOR t.)	• Broken agitator blade in second reactor	Replaced agitator shaft and impeller under warranty	Corrected (Period 2)
	• Broken shaft on reactor pump due to piece of rubber lining from agitator blade caught in impeller	Replace	Scheduled (1-2/76)
	• Failure of isolation valves on reactor pumps	Overhaul	Scheduled (1-2/76)
	INSTRUMENTATION:		
	• Erosion and plugging of flow-through pH unit	Replaced with immersion type probe with sonic cleaner	Corrected (Period 1)
FILTER/ THICKENER	EQUIPMENT:		
	• Erosion of fiberglass scraper	Replaced with SS under warranty	Corrected (Period 1)
	• Erosion of plastic bridge valve due to solids carried through cloth holes	Instructed operators to shut down filter and repair holes immediately	Corrected (Period 1)
	• Loosening of cloth retaining ropes out of caulking strips	Designed and installed new temporary caulking strips and reduced blower pressure	Corrected (Period 2)
	• Loss of vacuum due to cracks in internal fiberglass trunnion tubes in filter	Patched cracks--new caulking strip reduced stress on internals when installing cloth	Corrected (Period 2)
	• Erosion/cracking of fiberglass rocker arm on tub agitator	Reinforced with SS	Corrected (Period 1)
	• Insufficient agitation in filter tub to suspend sand-like particles and grit	Occasionally wash tub--(new agitation system to be installed)	Scheduled (1-2/76)
	• Plugging of thickener underflow lines	Installed flexible lines and new back-flushing provisions	Corrected (Period 1)
	• Deterioration of sections of lining in thickener and thickener hold tank	Patch and replace lining under warranty	Scheduled (1-2/76)

EQUIPMENT/INSTRUMENTATION STARTUP AND MAINTENANCE PROBLEMS

Scrubber	Problem	Action	Status
FILTER/ THICKENER (cont.)	INSTRUMENTATION:		
	• Poor level transmitter reliability in thickener hold tank	Replaced	Corrected (Period 2)
SODA ASH	EQUIPMENT:		
	• Clogging of dry feeder gate with lumps of soda ash	Replace feed control to allow higher gate position	Scheduled (1-2/76)
	• Failure of circuitry in heat tracing on piping	Replace	Scheduled (1-2/76)
	INSTRUMENTATION:		
	• Poor reliability of feed control system for soda ash liquor	Replace	Scheduled (1-2/76)
GENERAL	EQUIPMENT:		
	• Water freeze damage to pump seal water rotameters	Replaced and adjusted operating procedures	--

GLOSSARY

Active Sodium - Sodium associated with anions involved in SO_2 absorption reactions and includes sulfite, bisulfite, hydroxide and carbonate/bicarbonate. Total active sodium concentration is calculated as follows:

$$[Na^+]_{active} = 2 \times ([Na_2SO_3] + [Na_2CO_3]) + [NaHSO_3] + [NaOH] + [NaHCO_3]$$

Active Sodium Capacity - The equivalent amount of SO_2 which can be theoretically absorbed by the active sodium. Active sodium capacity is defined by:

$$[Na^+]_{active} \text{ capacity} = [Na_2SO_3] + 2 \times [Na_2CO_3] + [NaOH] + [NaHCO_3]$$

Calcium Utilization - The percentage of the calcium in the lime or limestone which is present in the solid product as a calcium-sulfur salt. Calcium utilization is defined as:

$$\text{Calcium Utilization} = \frac{\text{mols } (CaSO_3 + CaSO_4) \text{ generated}}{\text{mol Ca fed}} \times 100\%$$

Concentrated Dual Alkali Modes - Modes of operation of the dual alkali process in which the active sodium concentrations are greater than 0.15M active sodium.

Dilute Dual Alkali Modes - Modes of operation of the dual alkali process in which the active sodium concentration is less than or equal to 0.15M active sodium.

Sulfate Formation - The oxidation of sulfite. The level of sulfate formation relative to SO_2 absorption is given by:

$$\text{Sulfate Formation} = \frac{\text{mols } SO_3^{=} \text{ oxidized}}{\text{mol } SO_2 \text{ removed}} = 100\%$$

Sulfate Precipitation - The formation of $CaSO_4 \cdot XH_2O$ in soluble solids. The level of sulfate precipitation in the overall scheme is given by the ratio of calcium sulfate to the total calcium-sulfur salts produced:

$$\text{Sulfate Precipitation} = \frac{\text{mols } CaSO_4}{\text{mol } CaSO_X}$$

TOS--Total Oxidizable Sulfur - Equivalent to the sum of all sulfite and bisulfite species.

REFERENCES

1. LaMantia, C.R., Lunt, R.R., and Shah, I.S., "Dual Alkali Process for SO_2 Control," presented at Sixty-Sixth Annual Meeting, American Institute of Chemical Engineers (November 15, 1973), Paper No. 25c.

2. LaMantia, C.R., Lunt, R.R., Oberholtzer, J.E., Field, E.L., and Kaplan, N., "EPA-ADL Dual Alkali Program--Interim Results," presented at EPA Flue Gas Desulfurization Symposium, Atlanta (November 4-7, 1974).

3. Lunt, R.R., Rush, R.E., Frank, T.E., LaMantia, C.R., "Startup and Operation of the CEA/ADL Dual Alkali Process at Gulf Power/Southern Services," presented at The American Institute of Chemical Engineers, November 16-20, 1975.

4. Kaplan, N., "An Overview of Dual Alkali Processes for Flue Gas Desulfurization," EPA Flue Gas Desulfurization Symposium, Atlanta (November 1974).

APPLICABLE CONVERSION FACTORS

ENGLISH TO METRIC UNITS

British	Metric
5/9 (°F-32)	°C
1 ft	0.3048 meter
1 ft^2	0.0929 $meters^2$
1 ft^3	0.0283 $meters^3$
1 grain	0.0648 gram
1 in.	2.54 centimeters
1 $in.^2$	6.452 $centimeters^2$
1 $in.^3$	16.39 $centimeters^3$
1 lb (avoir.)	0.4536 kilogram
1 ton (long)	1.0160 metric tons
1 ton (short)	0.9072 metric tons
1 gal.	3.7853 liters
1 Btu	252 calories

THE FMC CONCENTRATED DOUBLE-ALKALI PROCESS

L. Karl Legatski, Karl E. Johnson, and Lyon Y. Lee

FMC Corporation
799 Roosevelt Road
Glen Ellyn, Illinois

ABSTRACT

FMC Corporation first developed its Concentrated Double-Alkali Process during the 1960's in response to internal needs. The system has since been commercialized by the Company's Environmental Equipment Division. The first full scale system was installed in one of the Company's own plants in 1971. Since then, it has been tested on a prototype scale (≈ 1 Mw) in ten different applications. A larger (3 Mw) demonstration plant has been in continuous operation for over one year, and the largest double-alkali system in the country was brought on stream in October, 1975.

FMC's development strategy and operating experience for these installations are discussed, and the economic and technical advantages of the system are enumerated. The system has demonstrated outstanding reliability and operability together with capital and operating costs equal to or less than lime/limestone systems.

Preceding page blank

INTRODUCTION

The FMC Concentrated Double-Alkali Process was initially developed by the Company's Industrial Chemical Division (ICD) in response to the need for a reliable flue gas desulfurization system for its own plants.

As early as 1956, ICD experimented with lime and limestone scrubbers and encountered many of the now well-known problems inherent in these processes. This early work led to a program to develop a throwaway process that could meet the following criteria:

1. High availability;
2. Easy to operate - within the capabilities of existing operating personnel; and
3. Capital and operating cost competitive with lime/limestone processes.

The FMC Concentrated Double-Alkali Process was the result of this program. It was first utilized as a large-scale prototype (about 30 Mw equivalent) in the Company's Modesto Chemical Plant in 1971. Responsibility for commercialization of the process for fossil fuel fired boilers was subsequently transferred to the Environmental Equipment Division. Since that time, the process has been tested and refined on a pilot plant scale and in several different commercial applications. A 3 Mw prototype has had 94% availability for the initial year of operation, and a 50 Mw equivalent commercial system started up in October, 1975 has also had a high availability since that time.

In late 1975, FMC received a patent on the Concentrated Double-Alkali Process covering what are believed to be the most feasible operating conditions of the process. FMC believes that its process has been sufficiently demonstrated in large scale installations, and it is aggressively pursuing large scale utility and industrial applications. The capital and operating costs compare very favorably to lime and limestone systems, and the reliability and ease of operation are demonstrated facts.

PROCESS DESCRIPTION

Figure 1 is a process schematic for the FMC Concentrated Double-Alkali Process. Essentially, the process entails a sodium based scrubbing loop in which sulfur dioxide is collected according to the reaction:

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3$$

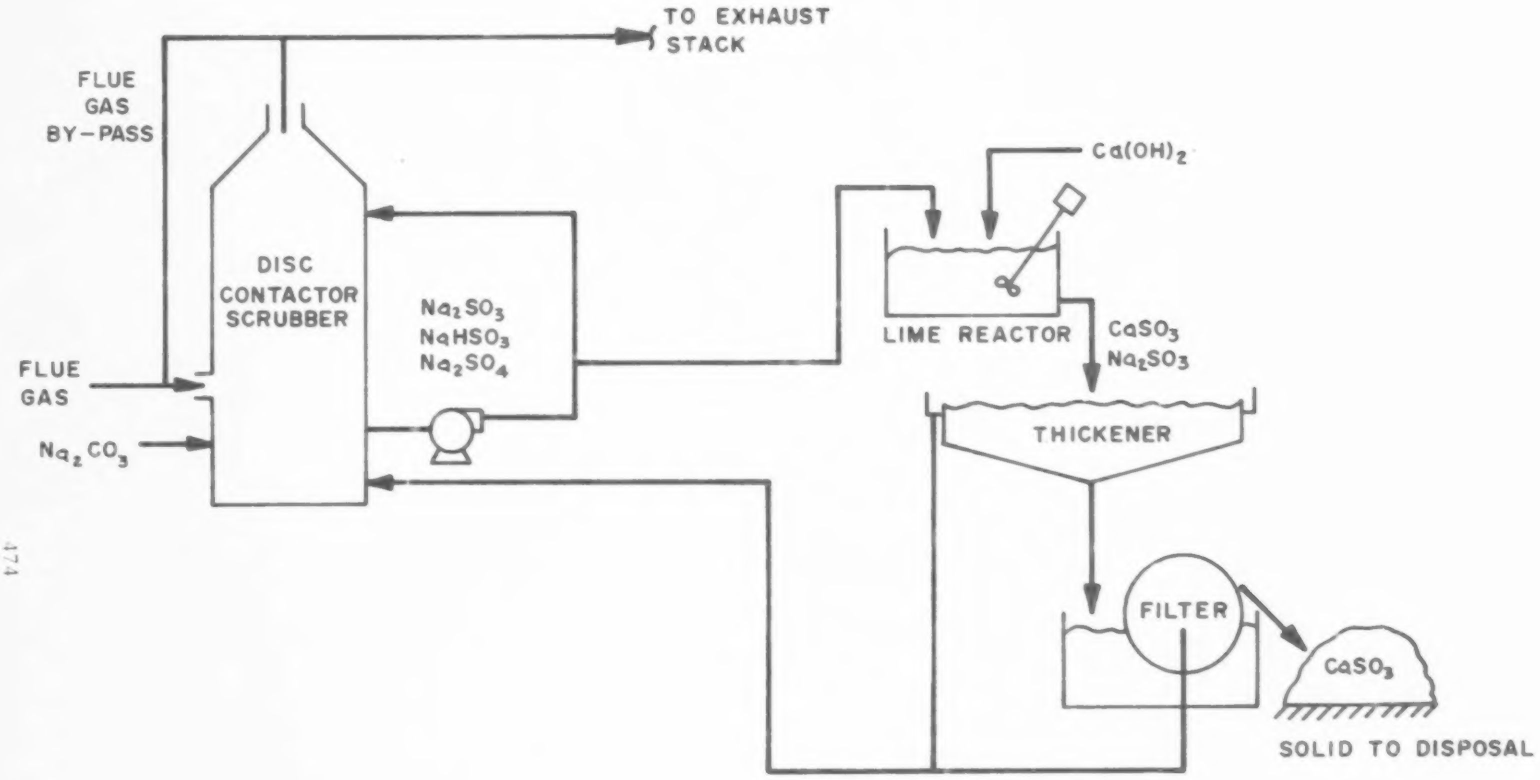
and subsequently precipitated with calcium in a separate loop according to the reaction:

$$2NaHSO_3 + Ca(OH)_2 \rightarrow CaSO_3 + Na_2SO_3 + 2H_2O$$

The principal advantages of the process in comparison to other types of throwaway processes are:

1. No scale in the scrubber;
2. Ease of control of the process;
3. Superior chemical utilization;
4. Improved solid waste properties; and
5. Ability to control sulfur dioxide and flyash concurrently.

One of the concerns in the double-alkali process is the oxidation of the sulfite ion in the scrubbing solution to sulfate. In contrast to the sulfite ion, the sulfate ion is not readily precipitated with the addition of calcium hydroxide and the sulfate ion therefore must be purged from the process at a rate equivalent to its net formation rate in the process. The loss of sodium compounds from the process is in the form of entrained liquor associated with the calcium sulfite and flyash insolubles. This loss is a problem in that alkaline sodium chemicals are more expensive than alkaline calcium chemicals and the soluble salts leaving the process may pose a potential water pollution problem. However, in comparison to lime and limestone processes, this is not a disadvantage in that the high magnesium limes that have been successful in lime scrubbing systems often produce a waste for disposal with a solubles content greater than or equal to a double-alkali process. Additionally, use of FMC's Concentrated Double-Alkali Process results in virtually stoichiometric utilization of chemicals and produces a mechanically stable filter cake of low permeability. The filter cake need not be fixed, which results in total operating costs substantially less than those of conventional lime/limestone systems.


TO EXHAUST STACK
FLUE GAS BY-PASS
DISC CONTACTOR SCRUBBER
FLUE GAS
Na_2CO_3
Na_2SO_3
$NaHSO_3$
Na_2SO_4
$Ca(OH)_2$
LIME REACTOR
$CaSO_3$
Na_2SO_3
THICKENER
FILTER
$CaSO_3$
SOLID TO DISPOSAL

SCRUBBING

$SO_2 + Na_2SO_3 + H_2O \rightarrow 2NaHSO_3$

REGENERATION

$Ca(OH)_2 + 2NaHSO_3 \rightarrow CaSO_3\downarrow + Na_2SO_3 + 2H_2O$

Figure 1. Schematic of FMC's Concentrated Double-Alkali Process.

In developing its double-alkali process, FMC was guided by the following objectives:

1. Develop the scrubbing solution composition and scrubber design that would minimize the net formation rate of soluble sulfate ion;
2. Utilize a scrubbing solution and regeneration solution chemistry that would be easy to control within the constraints of existing process control technology and be responsive to large and rapid changes in boiler operating conditions;
3. Develop the scrubbing chemistry that would minimize liquid flow rates and liquid holdup requirements in order to minimize the capital and operating costs of the facility; and
4. Develop a regeneration system that would yield a solid waste product physically and chemically acceptable for landfill disposal while minimizing the size and complexity of the equipment required for liquid-solid separation.

To accomplish these objectives, FMC has studied the chemical mechanisms involved in the process, the unit operations, the process controls and the materials of construction. The rate of soluble sulfate formation has been characterized as a function of the solution composition, flue gas composition, and scrubber design. Scrubber design has been optimized in terms of scrubber efficiency, pressure drop, and mist elimination capabilities. The chemistry of the lime regeneration step has been studied to maximize chemical utilization and minimize equipment size requirements. Thickener settling rates have been studied under a wide variety of operating conditions and vacuum filter performance has been investigated in terms of alternative filter design features and filter cake washing characteristics. The filter cake product has been evaluated chemically and physically on a laboratory and prototype scale for acceptability as a landfill. Materials of construction have been and continue to be routinely evaluated for acceptability under the various chemical and physical conditions experienced in the process. A proprietary control system for the process has been designed and demonstrated by virtually thousands of hours of trouble free operation.

The following sections briefly describe the state of FMC's development efforts in the various areas of process technology.

Scrubber Performance

The original type of scrubber utilized by FMC for sulfur dioxide control was its proprietary Dual Throat Venturi Scrubber. It was felt that a scrubber capable of simultaneous flyash and sulfur dioxide collection in a single contacting stage would offer significant advantages over the scrubbers utilized with lime and limestone processes which required flyash collection in a venturi followed by sulfur dioxide collection in an absorber. All of FMC's early pilot plant experience was with a venturi type scrubber, and the simultaneous collection capability was adequately demonstrated.

For stoker fired boilers, the flyash concentration could typically be reduced to 0.1 lbs/MM Btu simultaneously with sulfur dioxide collection efficiencies of 90% at scrubber pressure drops of approximately 10" w.g. However, the venturi type scrubber cannot be successfully operated at pressure drops significantly less than 10" w.g. without the risk of deterioration in collection efficiency. Furthermore, the tendency of utilities in recent years to utilize high efficiency electrostatic precipitators to comply with particulate regulations has decreased the potential market for simultaneous flyash and sulfur dioxide removal systems.

In 1973, FMC developed and tested its proprietary Disc Contactor Scrubber. This scrubber, which is essentially a baffle type scrubber, is designed to allow sulfur dioxide collection efficiencies in excess of 90% at a relatively low pressure drop without the use of spray nozzles. The original prototype was an 800 cfm unit that was tested on a boiler operated by Colorado Public Service Company in Boulder, Colorado. Subsequently, a 2000 acfm prototype unit was built and operated in conjunction with the trailer mounted pilot plant at St. Joe Minerals Corporation. The unit demonstrated an average collection efficiency of 94.9% operating on a pulverized coal boiler burning 2% sulfur coal.

While particulate removal was not a requirement of the scrubber, particulate removal efficiencies were also determined. When operating a scrubbing system with a scrubbing solution containing substantial quantities of dissolved or suspended solids, the possibility of the particulate emissions being adversely affected by entrainment of the scrubbing liquor is a potential concern. However, in a dissolved salt system, it is possible to utilize mesh type mist eliminators which provide excellent entrainment removal. This was demonstrated by tests where the scrubber inlet grain loadings, already below 0.01 gr/scf, were further reduced by the scrubber.

Furthermore, during a test where the precipitator was malfunctioning and was out of compliance with the 0.02 gr/scf guarantee, the Disc Contactor Scrubber was able to reduce the grain loading to within the specified level.

An additional test was conducted to determine the actual extent of entrainment from the scrubbing solution because the high sodium sulfate concentrations utilized in a concentrated double-alkali process may adversely affect total sulfate emissions. The results of these tests indicated that the entrainment of sodium sulfate was negligible.

With either type of scrubber, the Dual Throat Scrubber or the Disc Contactor Scrubber, FMC utilizes a scrubbing solution with a pH of approximately 6.5. If the pH is above 7, carbon dioxide absorption becomes significant and can lead to calcium carbonate scaling. A scrubbing solution pH below 6 is likewise avoided because the vapor pressure of sulfur dioxide increases dramatically for concentrated systems and can lead to equilibrium limited scrubbing conditions where outlet concentrations below 200 ppm are desired. This fact is demonstrated in Figure 2 in which the sulfur dioxide vapor pressure is plotted as a function of pH for a solution temperature of 130°F, a typical saturation temperature for a boiler flue gas stream. The vapor pressure of sulfur dioxide over dilute solutions is not significant, as can be seen from the line in the figure calculated from the Henry's Constant for sulfur dioxide and water and the first and second ionization constants for sulfurous acid for a solution with a total oxidizable sulfur (TOS) level of 0.03 gram moles/liter and an ionic strength of 0.3 gram moles/liter. However, for concentrated solutions, the sulfur dioxide vapor pressure at a pH of 6 is approximately 100 ppm and increases rapidly with decreasing pH.

For these reasons, FMC prefers to operate the scrubbing side of its process in the pH range of 6 to 7, and preferably at a set point of 6.5 where the sulfite-bisulfite system is highly buffered. The highly buffered system is preferred because the scrubbing solution is able to adapt very well to rapid changes in flue gas inlet conditions. This is in contrast to scrubbing systems utilizing a pH of 6 or less which can be very sensitive to rapid changes in flue gas inlet concentrations.

Sulfate Formation Rate

Superficially, the concept of the double-alkali process is quite straightforward. Sulfur dioxide is absorbed with a sodium sulfite solution, and the sodium sulfite solution is regenerated by precipitating the sulfur dioxide as calcium sulfite. The principal difficulty with the process is that sodium sulfite reacts very rapidly with oxygen to form sodium

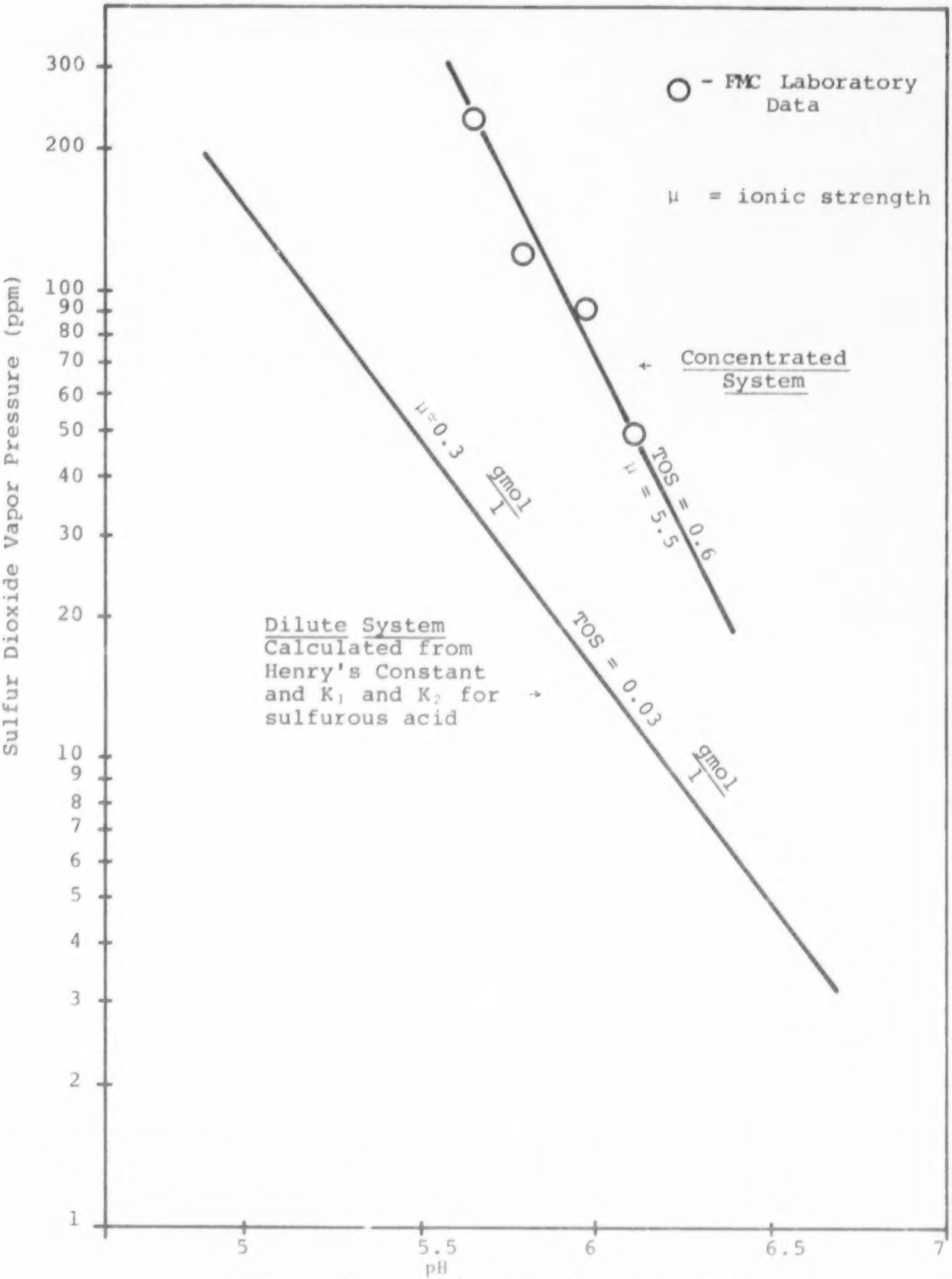


Figure 2. Sulfur dioxide vapor pressure.

sulfate, and the sodium sulfate cannot be readily regenerated into an active sodium form by reaction with lime. One of the principal areas of diversity in double-alkali process technology is the subject of sulfate formation and sulfate removal from the process.

FMC has found the effects of flue gas and solution composition on the rate of sulfate formation to be significant. A major part of FMC's development work has been to evaluate the effects of these parameters on the net formation rate of soluble sulfate. FMC first observed that, everything else being equal, increased ionic strength, principally in the form of increased sulfate concentration, reduced the net soluble sulfate formation rate. The results obtained during the pilot plant program indicated that a high ionic strength double-alkali process can be expected to produce a net soluble sulfate formation rate on stoker fired boilers operating with high excess air and flyash present equivalent to less than 10%, but probably not less than 5%, of the sulfur dioxide collected when operating with 2.5% or greater sulfur coal.

The effect of oxygen concentration on net soluble sulfate formation rate was also investigated. The results clearly indicated that with a pulverized coal boiler operating with 4 to 7% oxygen in the flue gas stream, the equivalent of approximately 30 to 60 ppm of sulfur dioxide will leave the process as sodium sulfate. While FMC has never been able to isolate any catalytic effect on oxidation caused by the presence of flyash, sulfate formation rates have generally been higher for stoker fired boilers with flyash present than with pulverized coal fired boilers with no flyash present. However, no data are available for low oxygen concentrations on stoker fired boilers and very little exists for high oxygen concentrations on PC fired boilers.

Inlet sulfur dioxide concentration has not been found to influence soluble sulfate formation. The sulfur dioxide concentrations encountered in pilot operations ranged from 400 to 8000 ppm. From these experiments, no correlation between net soluble sulfate formation rate and inlet sulfur dioxide concentration could be found.

Lime Reactor Performance

For the regeneration system, FMC utilizes the concept of regenerating the sodium bisulfite by reaction with lime according to the following reaction:

$$Ca(OH)_2 + 2NaHSO_3 \rightarrow Na_2SO_3 + CaSO_3 + 2H_2O$$

To accomplish this reaction, FMC utilizes a low residence time continuous stirred tank reactor. The lime reactor is controlled at a pH of about 8.5, which is effectively the titrametric endpoint for sodium bisulfite.

It is possible to precipitate additional sulfur values at elevated pH's according to the following reaction:

$$Ca(OH)_2 + Na_2SO_3 \rightarrow CaSO_3 + 2NaOH$$

However, it has been FMC's experience that it is difficult to prevent excess lime addition at pH's in excess of about 10. The responsiveness of the pH control system is excellent in the presence of bisulfite ion, but is relatively poor at the higher pH's existing in the lime-sulfite region. If a high pH set point is used, changes in scrubber operating conditions which lead to changes in the flow rate to regeneration can lead to the addition of excess lime due to the poor responsiveness of the pH control system. Excess lime leads to poor chemical utilizations and also results in poor filter cake quality. Furthermore, substantial precipitation of sulfite leads to increased calcium solubility in the system and potential supersaturation. The only advantage of operating the regeneration system at a high pH is the potential for minimizing the flow rate to regeneration by precipitation of additional sulfur values from the sodium sulfite solution.

FMC prefers to utilize a relatively high bisulfite concentration (typically 0.3 M) in the scrubbing solution which provides a flow rate to regeneration equal to or less than that obtainable in more dilute systems that attempt both to regenerate the bisulfite and precipitate additional sulfite by reaction with the sodium sulfite. The virtue of operating the lime reactor at the bisulfite endpoint is further demonstrated by the consistently high lime utilizations achieved by FMC. Independent analyses of the filter cake conducted by Firestone Tire and Rubber Company, General Motors Corporation, and St. Joe Minerals Corporation all indicate free lime concentrations in the filter cake of less than 1% on a dry basis.

The concentrated double-alkali process has the additional virtue of "built-in" softening due to the high dissolved sulfite concentrations always present in the scrubbing solution. This results in a dissolved calcium concentration an order of magnitude below its saturation level. Consequently, calcium sulfate scaling in the FMC process is virtually impossible.

In designing its lime reactor, FMC determined that for operation in the pH range of 8 to 10, residence times of only a few minutes were sufficient with reactor temperatures in excess of 120°F. At temperatures below 120°F (which may exist for applications such as acid plants) steam heat has been found to be an effective means of elevating the temperature to a level that will support the lime reaction.

Liquid-Solid Separation

FMC's Environmental Equipment Division is a leading supplier of thickening and clarification equipment for water treatment applications. The design of the thickening and clarification equipment for concentrated double-alkali processes is dependent upon many variables including the quality of the lime, the composition of the scrubbing solution, and the reactor temperature. FMC has experimented with rotary vacuum filters produced by several different manufacturers and prefers a knife discharge with an air puff to a belt filter because it requires less maintenance and produces a drier cake.

Vacuum filter performance results obtained from one installation recently are typical. These show an average percent solids in the filter cake of 61.3% for a two month period. The filter cake washing efficiency was approximately 70%. This was accomplished with a wash rate of from 2 to 3 displacement washes, reducing the soluble content of the filter cake on a wet basis to from 2 to 3%.

Landfill Studies

As a regular part of an ongoing development program, FMC has evaluated the filter cake product produced from this process for acceptability as landfill. These tests have been underway in laboratory and prototype demonstrations since 1973 under the direction of Dr. Raymond Krizek, Professor of Soil Mechanics, at Northwestern University. The tests originally concentrated on the physical properties of the material to determine its handleability, transportability, and bearing strength as a landfill material. It was found in the laboratory tests, and later confirmed in pilot plant demonstrations, that the material was amenable to transport on conveyor belts, could be loaded into trucks and subsequently moved, and showed a surprisingly good bearing strength.

Further tests were conducted to determine the leachability of the material. These tests showed that, while the soluble salts contained in the filter cake could be removed in shake tests, the material itself was highly impermeable, and proper management of a landfill could prevent displacement of the soluble salts by rain water.

A prototype landfill was constructed in conjunction with the Firestone Demonstration Plant. The landfill was designed according to good engineering practice, with appropriate test wells and reference wells plus provision for periodic core samples. These tests have been in progress over one year and, at this time, no contribution of soluble material to the ground water has been observed. Furthermore, the physical characteristics of the landfill have been extremely encouraging. The solid material is distributed in the pit area by a front end loader which maneuvers in the area without significant difficulty. Photographs of the filter cake and the landfill area are shown in Figures 3 and 4 and serve to demonstrate the favorable overall characteristics of the material.

FMC recognizes that some states may require fixation and/or the use of pond linings for disposal of the filter cake material, but will endeavor to prove that the material is physically and chemically acceptable for disposal in a non-chemical landfill without the use of chemical additives. There is no evidence to date that contradicts the above supposition, and the additional tests being conducted should confirm or refute FMC's position. In the worst case, a liner may be necessary, and initial cost estimates indicate that an appropriate lining material could be provided for less than $1.00/ton of disposal material. The filter cake produced by FMC's process should never have to be fixed to provide mechanical stability.

Water Balance

The FMC process is designed to dispose of all of the water that enters the process as either evaporation or entrained moisture on the filter cake. The degree of difficulty in achieving this goal depends on the sulfur dioxide concentration and the temperature of the flue gas stream. Higher sulfur dioxide concentrations require more lime slaking water and more filter wash water. Most of the water that leaves the process is in the form of evaporation, and the evaporation loss is directly related to the inlet flue gas temperature. Thus, a flue gas at 400°F containing 1500 ppm sulfur dioxide does not pose a water balance problem, while a flue gas stream at 280°F containing 3000 ppm sulfur dioxide will have a very tight water balance. For example, in the latter case, two displacement washes on the filter cake may consume 50% of the total makeup water requirement and lime slaking water may consume an additional 30 to 40%. The washing and lime slaking requirements, therefore, may consume virtually the entire water makeup requirement of the process.

The only uses of water in the process other than the filter washing and lime slaking water requirements are the pump seal water, soda ash solution makeup, and mist eliminator washing requirements.

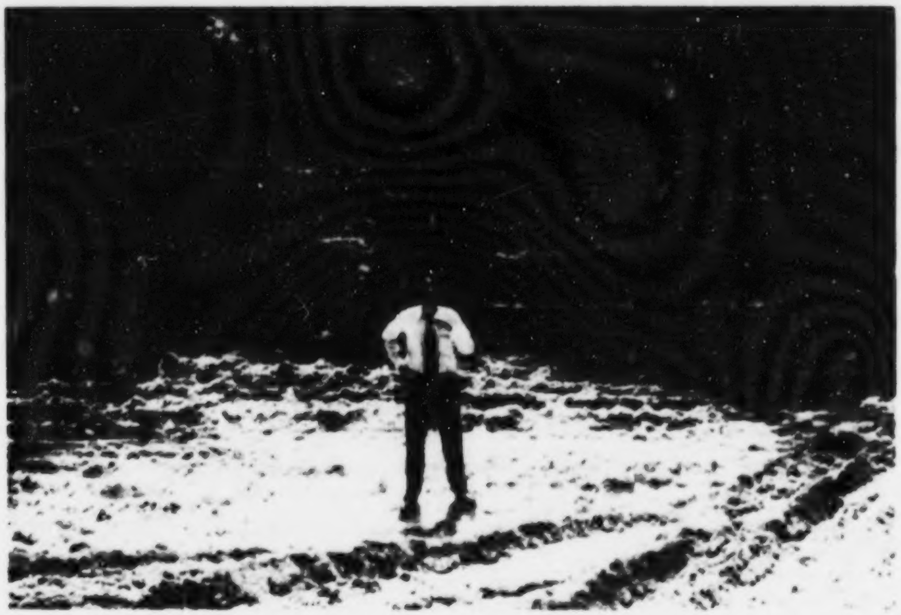


Figure 3. Landfill area at Firestone (note dozer tracks in right foreground).

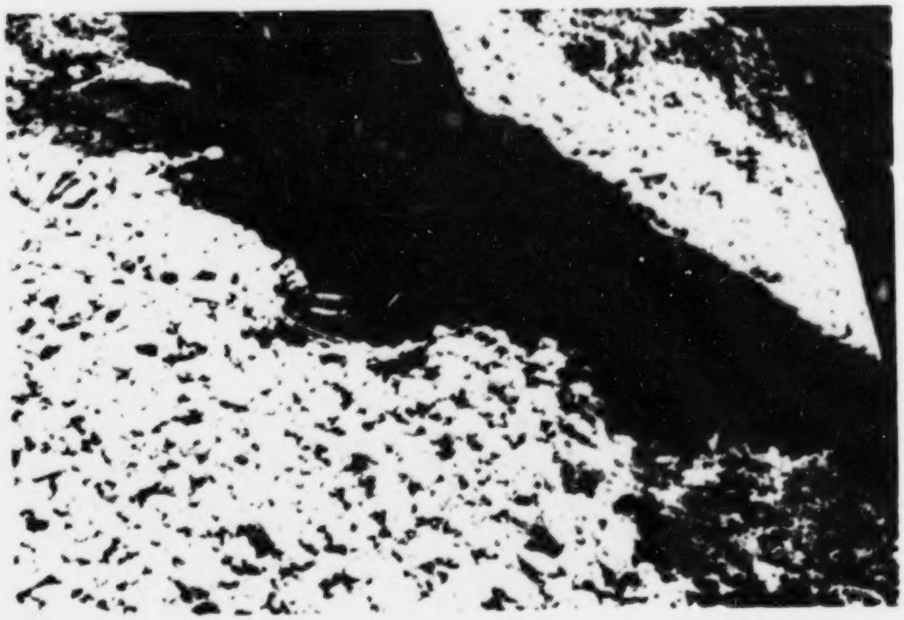


Figure 4. Climbing a recently dumped three-foot deep pile of waste material demonstrates physical stability.

OPERATING EXPERIENCE

FMC's operating experience can be summarized in five different applications or areas of experience:

1. The original installation at FMC's Modesto Chemical Plant;
2. Subsequent pilot plant work;
3. The Firestone Demonstration Plant;
4. The Caterpillar/Mossville installation; and
5. FMC's Green River Plant installation.

The most important facts concerning these projects are discussed below.

Modesto Chemical Plant

In 1956, the Modesto Chemical Plant of FMC's Industrial Chemical Division began work on an evaluation of the various possible processes for sulfur dioxide removal from flue gases. The Modesto Chemical Plant produces barium and strontium chemicals, such as the oxides, chlorides, carbonates and nitrates, by reducing barium and strontium sulfate in a reduction kiln to produce the respective oxides, which are reacted with the appropriate acids or soda ash in other parts of the plant. The reduction operation produces relatively high sulfur dioxide concentrations fluctuating in the range of 4000 to 8000 ppm.

The Modesto staff, with assistance from the Industrial Chemical Division's Princeton Research Center, initially tested lime and limestone scrubbers, but encountered many of the now classical problems with these systems. They next piloted a sodium based scrubbing system using soda ash, since soda ash was manufactured by FMC and already in use in the plant in other areas. The soda ash system proved to be very effective, but disposal of the sodium sulfite and sulfate products posed a problem. FMC was ultimately able to negotiate a contract with a nearby paper mill to sell the sodium sulfite and sulfate for makeup to their paper process. Unfortunately, just as detailed design of this system was completed, the paper mill cancelled the arrangement. Thus, the Modesto staff was faced with finding a method of recovering the sodium values from the sodium based scrubbing system while producing an acceptable solid waste for disposal. In response to this necessity, they developed the concentrated double-alkali process. This process was, in fundamental areas, the same as the process sold commercially today by FMC. The double-alkali process was successfully demonstrated by the Modesto staff on a pilot scale, and subsequently installed and operated on a full scale in 1971. Figure 5 is a photograph of the Modesto operation.

This unit, which processes a flue gas stream of 30,000 acfm is equivalent to approximately 30 Mw on the chemical regeneration side of the process. While the unit is not completely representative of the design FMC would utilize today, the system has demonstrated an extremely high degree of reliability in an application experiencing severe fluctuations in the process input conditions. Since startup, the process has been available in excess of 95% of the time, the majority of the non-available time accounted for by kiln down time rather than scrubber down time.

Pilot Plants

Shortly after the startup of the Modesto facility, the Air Pollution Control Operation began an evaluation of the potential for commercialization of sodium scrubbing and double-alkali processes. The processes were found to offer significant operating advantages in coal-fired applications in comparison to the competing lime and limestone processes, and the capital and operating costs appeared to be very competitive. Further, it seemed that if sulfur dioxide could be adequately collected in a venturi type of scrubber using a concentrated sodium based scrubbing system, the combining of flyash and sulfur dioxide collection into one operation would offer significant capital cost advantages.

In the Spring of 1972, a pilot test program was performed at Modesto using a venturi scrubber on a slip stream from the kiln gas system. This was the first in a long series of pilot plant experiments with the double-alkali process. This program demonstrated that a venturi type scrubber, such as FMC's Dual Throat Scrubber, could successfully collect sulfur dioxide in excess of 90% efficiency using the concentrated double-alkali process. Inlet sulfur dioxide concentrations of from 4000 to 8000 ppm were routinely reduced to the 100-200 ppm level.

Based on this encouraging experience, a pilot plant of the regeneration part of the process was built to operate in conjunction with the pilot scale venturi scrubber. This unit was installed on a slip stream from a stoker fired boiler at FMC's

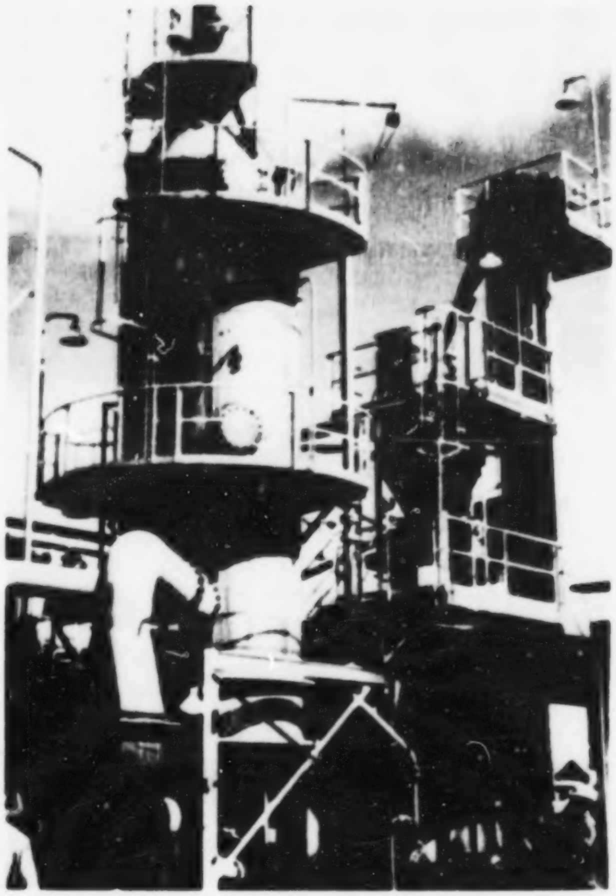


Figure 5. Concentrated Double-Alkali System installed at FMC's Modesto Chemical Plant.

Industrial Chemical Plant in South Charleston, West Virginia in late Spring, 1972.

The test at South Charleston lasted for approximately six months and consisted largely of experiments related to scrubber performance. The boiler was very old and seldom fired at full load, so the oxygen concentration was very high, typically 12 mole percent. This fact, combined with the relatively low inlet sulfur dioxide concentrations, which fluctuated from 400 to 1400 ppm, led to oxidation rates that represented a significant fraction of the sulfur dioxide collected. Thus, the majority of the effort at South Charleston was devoted to determining the effects of solution composition and scrubber operating conditions on the sulfur dioxide collection efficiency and the rate of soluble sulfate formation.

At the conclusion of the test program, it was decided that the process was technically feasible for application to coal fired boilers. The Air Pollution Control Operation was incorporated into the Environmental Equipment Division, and the decision was made to pursue the sulfur dioxide control market in the industrial boiler field. The pilot plant was installed in a 35 ft. van to provide efficient transportation to the facilities of potential customers. In the Spring of 1973, a contract was obtained from Caterpillar Tractor Co. for a two month demonstration of the process on a slip stream from a stoker fired boiler at the heating plant of Caterpillar's Mossville Engine Plant.

The pilot plant was operated on an 8 hour per day basis for two months by two FMC Engineers, and the results were completely satisfactory to Caterpillar and consistent with the prior pilot plant experience by FMC. As a result of this successful operation, Caterpillar Tractor Co. awarded FMC a contract in May 1973 for turnkey installation of the sulfur dioxide control system at the Mossville Engine Plant.

The pilot plant program was ultimately extended to ten different applications in which the process was applied for periods from a few weeks to several months. The range of the input variables experienced by the system was substantial:

SO_2 concentrations - 400 to 8000 ppm

O_2 concentrations - 5 to 12%

Temperature - 120 to 500°F

Particulate - Nil to 3.0 gr/sdcf

These varied inputs allowed FMC to develop a knowledge of the phenomena that govern sulfite oxidation and sodium consumption that is unsurpassed in the industry.

Firestone Demonstration Plant

In January of 1975, FMC completed the installation of a three Mw equivalent demonstration plant (Figures 6 and 7) for The Firestone Tire & Rubber Company in Pottstown, Pennsylvania. This installation, utilizing a 10,000 acfm Dual Throat Venturi Scrubber, is operating on a slip stream from a pulverized coal boiler that is presently burning 2% sulfur oil to meet particulate emission regulations in the State of Pennsylvania. While this unit is not of commercial scale and is not officially a commercial operation, the experience at Firestone demonstrates four very important points about the process.

First, the installation is of major significance because virtually all of the process data and relationships that were developed in the pilot plant program have been confirmed in this installation, which is four or five times as large as the trailer mounted prototype unit.

The second key feature of the project is the proven operability of the system. Two weeks after startup, the operation of the unit was turned over to three boiler house operators (one per shift) who have been responsible for the operation since that time. FMC provides one engineer on the day shift to perform chemical analyses and experiments related to the performance of the various unit operations, but he has virtually no responsibility for the routine operation of the system. Performance has been so good and the system has required so little operator attention, that a remote alarm system is now being installed to allow the operators to work their normal shift at the boiler control panel and inspect the demonstration plant area only once every two hours or whenever a common alarm sounds at the boiler panel indicating an upset in the process.

Third, the landfill study program discussed in a preceding section has demonstrated the handleability and impermeability of the filter cake waste material.

Finally, the availability of the operating system has been outstanding. For the entire first year of operation, including most of the start-up period, the overall availability has been 93.5%. This is particularly impressive in view of the lack of any spares in the system and the fact that the operation was often unattended on the night shift.

The various segments of the system that contributed to the downtime are itemized in Table 1. The single longest outage was due to blockage of the thickener underflow pipe by a plastic beaker that was unintentionally dropped into the thickener. The second longest outage was caused by improper maintenance of the recirculation pump seal, which led to separation of the pump lining from the pump housing. The

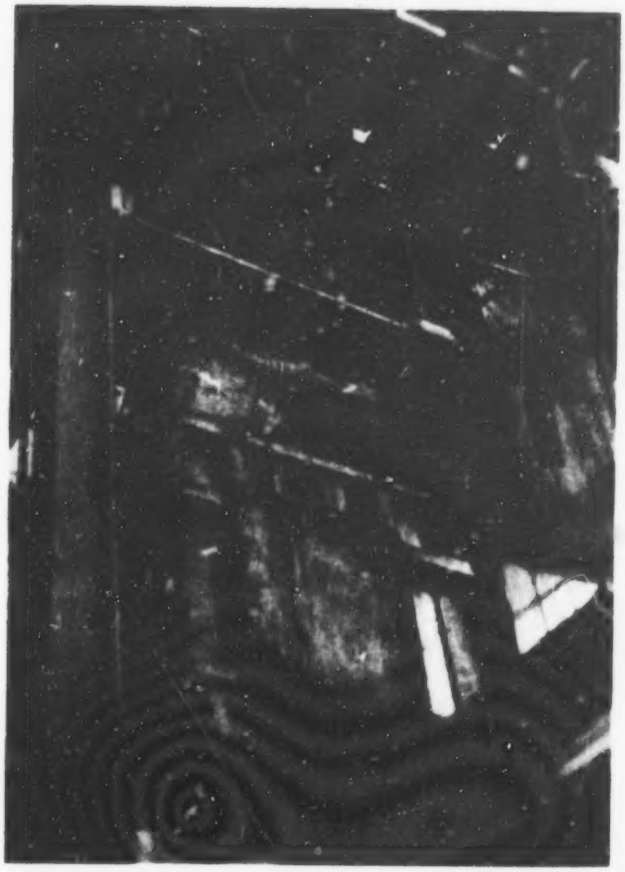


Figure 6. FMC Dual-Throat Scrubber and Cyclone in Firestone Demonstration Plant.

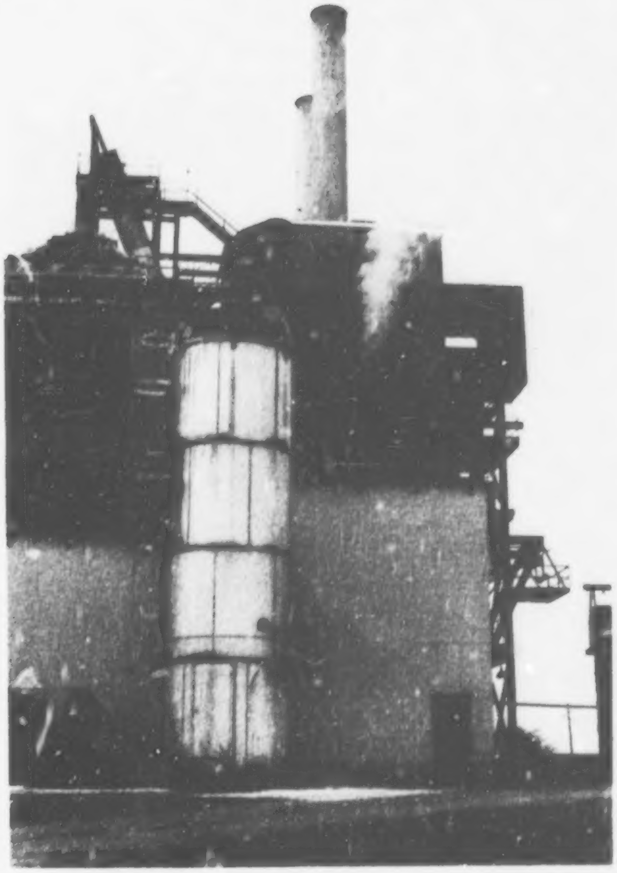


Figure 7. Lime storage tank and dumpster at Firestone Demonstration Plant.

Table 1. AVAILABILITY OF FIRESTONE DEMONSTRATION PLANT

	Item	Downtime % of Total Downtime	 % of Operating Year
1.	Thickener Pluggage	20.7	1.35
2.	Pumps	16.1	1.04
3.	Cake Conveyor	16.0	1.04
4.	Fan	15.4	1.00
5.	Lime Feeder	13.4	0.87
6.	Spray Nozzles	12.4	0.80
7.	Instrumentation	4.1	0.27
8.	Control Valves	1.9	0.13
		100.0	6.50

Total Annual Availability = 93.5%

Note: The Firestone Demonstration Plant has no installed spares.

filter cake conveyor problems generally related to the fact that an inappropriate conveyor was used, due to the unavailability of high quality equipment for the small size required. The fan outage was due to overheated bearings, which were in turn caused by shaft misalignment. The remaining outages were similarly due to problems unrelated to the process, and all of them could have been eliminated or substantially reduced by a normal commercial sparing philosophy.

Caterpillar Mossville Installation

FMC's largest double-alkali installation to date is the system for Caterpillar Tractor Co. at its Mossville Engine Plant outside of Peoria, Illinois. The unit is the largest double-alkali system in the country and is designed to collect sulfur dioxide and flyash from four stoker fired boilers totaling 460,000 lbs/hr of generating capacity. The operation of this unit commenced in October, 1975. Figures 8 and 9 illustrate portions of the system.

During startup and initial testing, the Caterpillar Mossville unit has demonstrated a high availability. At the present time, two of the boilers are in operation, and the system has not yet been tested at full load. Furthermore, one of these boilers is new and has been operated only intermittently because it is still being debugged. However, there has been only one occasion since startup in which a boiler has been shut down due to a problem with the FGD system, and the scrubbers have not been modified or maintained during the boiler outages. FMC expects the high availability achieved during startup to continue or improve during sustained boiler operations.

One of the key features of the Mossville installation is the simultaneous collection of flyash and sulfur dioxide. There are no mechanical collectors on these boilers, and the full flyash load is removed by the scrubbers. This fact, combined with the high excess air inherent in the stoker boilers, creates conditions that are far more technically difficult for a double-alkali process than a low excess air pulverized coal boiler with a high efficiency precipitator attached. At Mossville, most of the problems to date have been related to wear with the control valves, filter cloth, and conveyor belts. At this time, FMC feels that it has arrived at satisfactory solutions to all of these problems. In the long run, the concentrated double-alkali process offers the potential for reliable, simultaneous collection of sulfur dioxide and flyash in utility and industrial applications. This would result in substantial savings in capital and operating costs, and FMC feels that this will be one of the major advantages of double-alkali systems in the future.

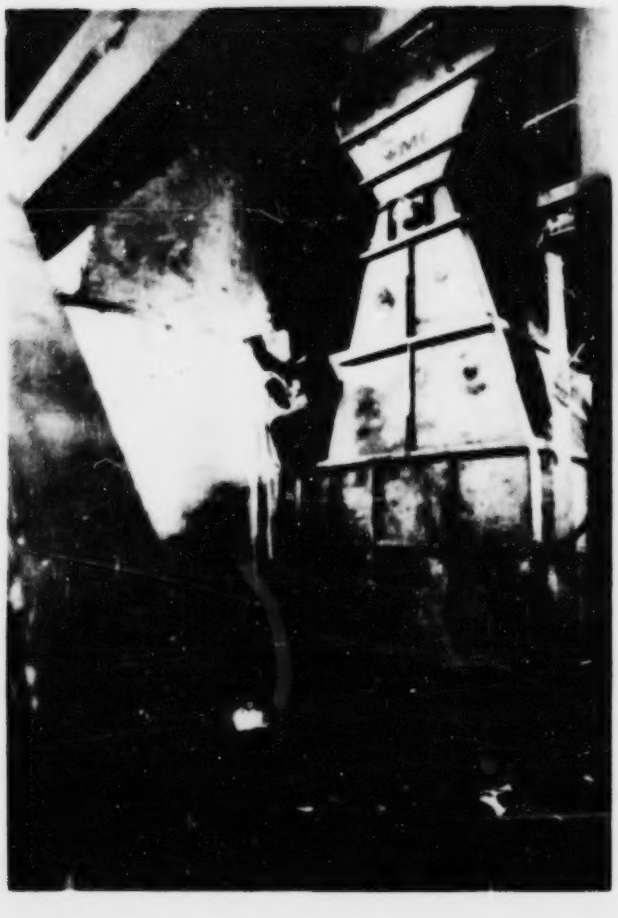


Figure 8. Dual-Throat Scrubber (80,000 ACFM) being installed on 150,000 lb./hr. boiler at Caterpillar Tractor Co.

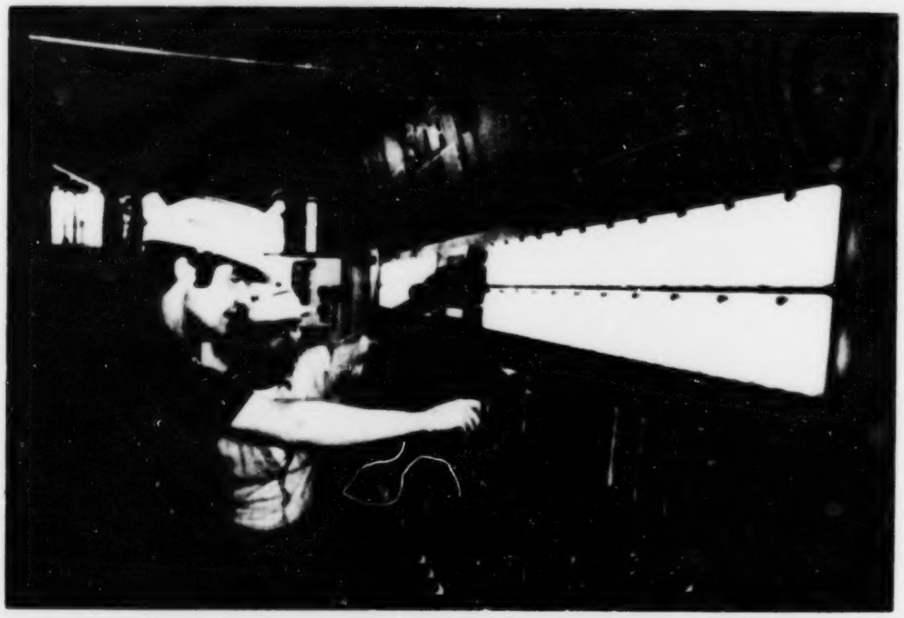


Figure 9. Control panel at Caterpillar Mossville installation.

Based on FMC's performance to date on the Mossville project, Caterpillar Tractor Co. has awarded FMC contracts for the design of complete SO_2 and particulate control systems for its East Peoria and Mapleton plants. Each system will have a heat input equal to approximately 100 Mw. The engineering design for these units has been completed and ultimate startup of the two systems is expected to be in 1977 and 1978, respectively.

FMC Green River

This Spring, FMC will start up its largest sodium scrubbing system which consists of two 330,000 acfm Disc Contactor Scrubbers on the new pulverized coal fired boilers now being installed at FMC's Green River Soda Ash Plant in Green River, Wyoming. The boilers are two 625,000 lbs/hr units followed by high efficiency electrostatic precipitators. The scrubbers are designed to remove 90% of the sulfur dioxide at a pressure drop of 5.5" w.g., and are equipped with a bypass reheat arrangement.

This represents the third different type of reheat arrangement used by FMC in commercial installations. The Caterpillar Mossville facility, due to spatial limitations, uses Dual Throat Scrubbers followed by steam tube reheaters operating on an ambient air stream injected into the scrubber discharge, followed in turn by an induced draft fan. The East Peoria and Mapleton installations for Caterpillar utilize forced draft fans on the scrubbers, followed by provision for steam tube air injection reheat. The use of steam tube air injection reheat was dictated by Caterpillar to reduce its dependency on fuels other than coal. Due to the low sulfur dioxide collection efficiency requirement at Green River, the scrubbers are designed to utilize bypass flue gas for reheat whenever possible and oil fired reheat when higher sulfur coal is burned or the boiler is at full load.

CAPITAL AND OPERATING COSTS

It is difficult to generalize on the capital and operating costs for a flue gas desulfurization system, because of the differences in design bases and input assumptions that can occur. Due to significant operating advantages of the double-alkali system, many people have arrived at the conclusion that it is inherently more expensive than lime/limestone. Closer examination of both systems indicates that this conclusion is not justified.

In capital cost, it is superficially obvious that double-alkali and lime/limestone processes must be comparable. In fact, the comparison in Table 2 indicates that double-alkali may be less expensive. Virtually every component in the double-alkali process has an equal or larger counterpart in a lime/limestone process. The only potential disadvantage of double-alkali is that it requires filters, but many lime/limestone systems now use filters.

In operating costs, the concentrated double-alkali system is generally less expensive in all areas. Chemical consumption is essentially equal to lime systems; limestone may be less depending on location.

Power requirements are less due to lower pressure drops and liquid rates. Water requirements are slightly less. Reheat is often less because the high efficiency of double-alkali makes bypass reheat possible. Maintenance and operating labor are obviously less in the highly buffered, soluble salt, concentrated double-alkali system. Concentrated double-alkali systems do not scale, and scrubbing with a solution instead of a slurry results in substantially less wear.

Finally, solid waste disposal is substantially less expensive. There is less material to begin with because of the superior chemical utilization and the low moisture content of the filter cake. Chemical fixation is not required for mechanical stability.

A typical example of estimated operating costs for a 300 Mw Midwestern utility burning 3.13% sulfur coal is given in Table 3.

Table 2. QUALITATIVE CAPITAL COST COMPARISON OF DOUBLE-ALKALI AND LIME/LIMESTONE SYSTEMS*

Equipment Item	Lime/Limestone	FMC Concentrated Double-Alkali
1. Scrubbers	-	Equal or smaller (less mass transfer units)
2. Fans	12 to 15" w.g. typical	Smaller (6" w.g. or less)
3. Pumps	L/G = 30 to 40 typical	Smaller (L/G = 10 gal/mcf)
4. Thickener	-	Equal
5. Filter	Less or Equal	-
6. Tankage	-	Less (low residence time)
7. Materials in	Less storage for limestone, but substantially more material used	About same as lime
8. Materials out	-	Substantially less

*No flyash removal included.

Table 3. ESTIMATED OPERATING COSTS FOR THE FMC CONCENTRATED DOUBLE-ALKALI SULFUR DIOXIDE REMOVAL SYSTEM[1]

Chemicals	Consumption	Unit Cost	¢/KWH
Lime (92% CaO)	0.047 lb/KWH	\$33/ton	0.0776
Soda Ash	0.0021 lb/KWH	\$64/ton	0.0067
Utilities			
Power	0.0137 KWH/KWH	1.75¢/KWH	0.0240
Water	0.078 gal/KWH	40¢/Mgal	0.0031
Fuel Oil[2]	0.0006 gal/KWH	35¢/gal	0.0210
Operating Labor & Supplies	1 man/shift + supplies	\$30,000/man year + \$50,000 supplies	0.0067
Maintenance	3% of Plant	0.03x\$21 MM	0.0240
Solid Waste Disposal	0.204 lb/KWH	\$4.00/ton	0.0408
TOTAL OPERATING COST (excluding capital charges)			0.2039

[1] 300 Mw unit operating at full load with 3.13% sulfur coal; 90% SO_2 collection; system designed for 10% over heat input using the maximum sulfur, minimum Btu coal expected.

[2] 50°F of total reheat: 25°F with oil, and 25°F with bypass flue gas.

CONCLUSIONS

In conclusion, there are substantial advantages to the FMC process in comparison to other systems. In comparison to other double-alkali systems, FMC offers the following:

1. FMC has more cumulative hours of operating experience in pilot plant, prototype, and commercial systems on more different applications than anyone else. Due to the breadth of its experience, FMC is in a position to specify performance standards and accept significant financial obligations as a consequence of potential nonperformance.

2. The sulfur dioxide collection capabilities of FMC's process and equipment are demonstrated facts. With sulfur dioxide control systems there is a potential risk that the scrubbing liquor entrainment will adversely affect the particulate and/or primary sulfate emissions. FMC has demonstrated that its Disc Contactor Scrubber does not adversely affect fine particulate or primary sulfate air emissions.

3. In terms of process considerations, FMC has demonstrated essentially 100% utilization of makeup chemicals (soda ash and lime) through thousands of hours of prototype and commercial operation. Scale prevention is inherent in the FMC process due to sulfite softening of the scrubbing liquor in the regeneration portion of the process. At no time in either its commercial or prototype experience has FMC experienced a scaling problem in its scrubber or elsewhere in its process. FMC has an excellent knowledge of the causes of sulfate formation and has demonstrated the capability to predict and inhibit the net sulfate formation rate it will experience by controlling the process operating conditions.

4. The principal objective of all of FMC's development work has been to develop a process that takes advantage of the inherent reliability of a sodium scrubbing process while striving to keep the process simple enough to compete in capital

and operating costs with lime and limestone systems. The inherent reliability of the process has been demonstrated in the Modesto Plant of FMC and the Firestone Demonstration Plant. The overall simplicity of the process is apparent from the process description contained in this section.

5. FMC's Double-Alkali Process is proprietary and is patented (U.S. Patent No. 3,911,084). FMC believes the operating conditions specified in the patent are optimal for operating a concentrated double-alkali process.

In comparison to lime/limestone processes, FMC believes it offers the following significant advantages:

1. Ability to achieve higher SO_2 removal efficiencies. This often allows the use of bypass reheat resulting in a substantial savings.
2. Sodium scrubbing processes are much more tolerant of process upsets and variations in gas flow and composition.
3. The reduced liquid-to-gas ratio (usually 10 gallons per 1000 acfm) results in smaller pumps and tanks.
4. The double-alkali process produces a drier and more mechanically stable filter cake.
5. The sodium scrubbing processes produce a lower outlet particulate loading due to no slurry in the scrubber. This allows the use of a mesh type mist eliminator without plugging.
6. The SO_2 collection efficiency is variable over wide limits by simply adjusting the chemistry. This can be significant in cases where a high SO_2 collection efficiency is not desirable or necessary.
7. The scrubbing solution is highly buffered so corrosion is reduced. This also results in a more operable and less sensitive system.
8. Calcium scaling does not occur. FMC has never shut down a scrubber because of scaling in an SO_2 application.

9. One of the most significant advantages of sodium scrubbing is the ability to collect SO_2 and flyash concurrently. This can result in substantial capital cost savings.

10. FMC's installations are being operated by boiler house personnel. All installations have had in excess of 90% availability, and the average has been about 95%. The demonstrated reliability and operability of FMC's installations is unsurpassed in the industry, to our knowledge.

FMC continues to develop and refine its basic process through an intensive research and development program in the following areas:

1. Improved scrubber design to reduce capital and operating costs;

2. Improved thickener design to reduce space requirements;

3. Improved filter cake washing to further reduce sodium losses; and

4. Continued landfill evaluation program to demonstrate the acceptability of disposal without further treatment.

ACKNOWLEDGEMENTS

FMC Corporation gratefully acknowledges the assistance of Messrs. Roman Zaharchuk and Gary W. Wamsley of The Firestone Tire & Rubber Company and Mr. David C. Dietrich of Caterpillar Tractor Co., all of whom made material contributions to the success of the projects discussed in this paper.

OPERATING EXPERIENCE WITH THE ZURN DOUBLE ALKALI FLUE GAS DESULFURIZATION PROCESS

P. M. Lewis

Zurn Industries, Inc.
Air Systems Division
Birmingham, Alabama

ABSTRACT

The first Zurn Double Alkali Flue Gas Desulfurization went on-line in September, 1974, at the Joliet Plant of Caterpillar Tractor Company, Joliet, Illinois. The system is scrubbing the flue gas from two coal fired boilers rated at 100,000 lb./hr. and 80,000 lb./hr. steam. These boilers produce steam for building heat and are in service about seven months a year. The system utilizes a dilute double alkali process with lime addition for regeneration of the spent scrubbing solution, and sodium carbonate addition for scrubber feed water softening. The calcium based sludge generated by the process is dewatered for disposal in a remote landfill.

This paper summarizes the process, and discusses operating experiences to date. Process reliability, equipment performance, and solutions to problems are discussed.

OPERATING EXPERIENCE WITH THE ZURN DOUBLE ALKALI FLUE GAS DESULFURIZATION PROCESS.

I. INTRODUCTION

Due to the uncertain future availability and cost of natural gas and fuel oil, Caterpillar Tractor Co. decided in 1970 to rely on coal for steam generation in two boilers at the Caterpillar Tractor Co. plant in Joliet, Illinois.

Basically, there are two methods for complying with sulfur oxide emission standards while burning coal. These are, removing the sulfur oxides from the flue gas before it is emitted to the atmosphere, and the use of low sulfur coal. The demand for low sulfur coal was increasing rapidly, and its future price and availability prospects were not favorable. The use of high sulfur coal and flue gas desulfurization appeared to be the most economical alternative that would provide an adequate fuel supply and ensure compliance with current regulatory requirements. In particular, after technical and economic evaluations, they selected the Zurn Double Alkali Flue Gas Desulfurization System.

It is the purpose of this paper to describe this system and to relate the operating experience at the Joliet, Illinois installation to date.

II. SYSTEM DESCRIPTION

The Zurn Double Alkali Process uses a dilute solution of sodium hydroxide to absorb the sulfur oxides from flue gas. Lime is used to regenerate the spent scrubbing solution to form sodium hydroxide and insoluble sulfur bearing calcium compounds which can be removed by mechanical means.

This double alkali process can be viewed as having three functionally different process stages. As shown in Figure 1, these process functions are gas scrubbing, chemical regeneration, and solids removal.

GAS SCRUBBING

The function of the gas scrubbing section is to intimately mix the flue gas containing sulfur dioxide with a dilute sodium hydroxide solution. This reacts with the sulfur dioxide gas and converts it to water soluble, sulfur containing compounds (sodium sulfite, sodium bisulfite, and sodium sulfate). The scrubber also removes a high percentage of any fly ash or other particulate which may be contained in the flue gas.

The reactions that take place in the scrubbing section of the system are shown below:

$$2\ NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O \qquad (1)$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \qquad (2)$$

$$Na_2SO_3 + 1/2O_2 \rightarrow Na_2SO_4 \qquad (3)$$

Sodium hydroxide and sulfur dioxide react to form sodium sulfite (reaction 1). The sodium sulfite then reacts with sulfur dioxide and water to form sodium bisulfite (reaction 2). Some of active sulfite is oxidized to sodium sulfate by excess oxygen in the combustion gases (reaction 3). After being mixed with the flue gas, the solution is withdrawn from the scrubber to an external mix tank for chemical regeneration of the spent scrubbing solution.

REGENERATION

The purpose of the chemical regeneration section is the conversion of the spent scrubbing solution to a usable form, and the separation of this solution from the sulfur containing solid compounds. In the mix tank, calcium hydroxide, from previously slaked quick lime, is added to the spent scrubber solution. The calcium hydroxide reacts with the soluble sodium salts to form insoluble calcium salts (calcium sulfite and calcium sulfate (or gypsum). This reaction is effected in two agitated process tanks and the resultant mixture is then pumped to a thickener where the calcium salts precipitate and are removed as a slurry. The regenerated sodium hydroxide solution overflows to a clarifier. The reactions in the regeneration section are:

$$NaHSO_3 + Ca(OH)_2 \rightarrow NaOH + CaSO_3 \downarrow + H_2O \qquad (4)$$

$$Na_2SO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_3 \downarrow \qquad (5)$$

$$Na_2SO_4 + Ca(OH)_2 \rightarrow 2NaOH + CaSO_4 \downarrow \qquad (6)$$

Sodium carbonate is added to the clarifier to further reduce the concentration of calcium by the precipitation of calcium carbonate (limestone). This step is referred to as "softening". The further precipitation of calcium in this step reduces the likelihood of calcium salts precipitating in the scrubbing section. The softening reaction is as follows:

$$Na_2CO_3 + Ca(OH)_2 \rightarrow 2NaOH + CaCO_3 \downarrow \qquad (7)$$

After the solution has passed thru the clarifier, it has been regenerated and is available for reuse in the scrubbing section to remove more sulfur dioxide. The insoluble calcium salts have been removed in separate process equipment specifically designed for solids precipitation and concentration.

LIQUID-SOLID SEPARATION

The underflow slurry from the thickener and clarifier is pumped to the solids removal section of the process for further dewatering. The function of the solids removal section is to recover the usable scrubbing solution contained in the thickened underflow slurry and to convert the slurry solids into a form which is easily handled and is acceptable for disposal. This is accomplished with rotary drum vacuum filters which convert the slurry into a cake containing approximately 50 to 60% by weight solids, with the remainder being water. These filters also have provisions for "washing" the cake to remove as much of the soluble sodium compounds as possible before disposal. The filter cake is composed of calcium sulfite, calcium sulfate (gypsum), calcium carbonate (limestone), and the particulate removed from the flue gas by the scrubber.

The filter cake discharges to an automated conveyor system which alternately fills waste product containers for off-site disposal as landfill. The filtrate separated from the cake by the vacuum filters is recycled back to the thickener to be reused in the system. Thus, the process operates in a "closed loop", which means there is no process solution discharged from the system other than that contained in the filter cake.

III. PERFORMANCE CHARACTERISTICS

DESIGN BASIS

The Zurn Industries, Inc. system installed at Caterpillar Tractor Co.'s Joliet, Illinois plant cleans the flue gas from the two stoker fired boilers which produce steam for building heat and process needs. These boilers are rated at 80,000 and 100,000 lbs. of steam per hour and each is equipped with a mechanical dust collector. Total coal firing rate to these boilers is 9.65 tons per hour and the process is designed to allow the boilers to operate within the allowable emission rate of 1.8 lbs. of SO_2 per million BTUs heat input. System design parameters are summarized below:

No. of Boilers	2
Firing Method	Stoker
Total Steam Capacity	180,000 Lbs./Hr.
Total Gas Volume	103,500 Cu. Ft./Min.
Coal Sulfur Content	4% As Received
SO_2 From Boilers	6.74 Lbs./mmBTUs
Allowable SO_2 Emission	1.8 Lbs./mmBTUs

The total process design, engineering, construction, and start-up of this system were accomplished by Zurn Industries, Inc. Design, construction, and start-up were completed on, or ahead of schedule. The chronology of system design and installation was as follows:

1. Feasibility investigations 15 Months
2. System design 12 Months
3. Equipment installation started February 1, 1974
4. Component check-out started August 12, 1974
5. First scrubber put in service September 25, 1974
6. Second scrubber put in service October 2, 1974

EQUIPMENT - PROCESS PERFORMANCE

The scrubbers, breeching, stack, and scrubber effluent piping are constructed of 316 stainless steel. The wet surfaces of the vacuum filters are 304 stainless steel. All other process piping and process tanks are carbon steel. The lime mix tanks are lined with a Urecal polyurethane coating.

System start-up and preliminary testing was accomplished from September 25 through December 16, 1974. The major reasons for downtime during this period were minor mechanical equipment problems typical of system start-ups, and the accumulation of unslaked lime and grit in the lime mix tanks. This accumulation was due to poor slaking characteristics of the delivered lime, high insolubles content, and some air-slaking of the lime in the storage silo. Accumulation rates were quite dramatic. Approximately 10 feet of sand, gravel, and unslaked lime were deposited in the primary lime mix tank in one ten day period. This problem was solved by the installation of a lime slaker which improved lime slaking and had provisions for separating sand and gravel from the lime before it entered the system. The slaker installation was accomplished during the scheduled Christmas 1974 shutdown.

Design revisions were completed and the system was put back on line January 23, 1975, and operated continuously until June 1, 1975. Process performance and reliability during this period was quite good. One of the operation highlights observed during the prolonged period of operation was good system control over a wide range of system loads. The scrubber pressure drop and pH were very stable, easily controlled, and were not adversely affected by rapid boiler load changes. No significant accumulation of solids was observed in the scrubbers which would impede their efficiency or operation. No measurable corrosion or erosion was observed in any of the process vessels, scrubbers, or piping.

Various mechanical problems with process auxiliaries were encountered during this period and some of these are illustrated below.

Chemical Feed System

The lime and soda ash are metered to the process from outside silos by screw conveyors. Two problems were encountered here. One was the reliability of components in the electronic control system for the conveyor drives and the other was due to formation of lumps inside the silos caused by moisture from rainwater leakage and condensation on the inside tank walls.

Both of these problems cause inaccurate feed rates which in turn caused undesirable deviations in system chemistry. The conveyor drive control systems were made operational by replacing faulty electronic components. Inaccurate feed rates due to lumps of material forming inside the silos was a continuing problem throughout this period of operation. A dry air purge system was installed on the lime and sodium carbonate storage silos to prevent material degradation and lump formation due to moisture. This has effectively prevented lump formation in the storage silos, and dry material flow from the silos has been reliable.

pH Probe Reliability

Accurate pH indication is essential to maintaining good system control, and preventing scaling and corrosion. The pH probes as initially installed were quite susceptible to coating and fouling which resulted in erroneous readings. Several methods were investigated to alleviate this problem and reliable pH monitoring was finally achieved by the installation of parallel probes with fresh water back flush and ultrasonic cleaning.

Sulfur Dioxide Monitoring

Stationary probes are installed in the scrubber inlet duct and in the stack to provide continuous monitoring of flue gas for SO_2 concentration. The SO_2 analyzer was not operational during initial system check-out due to shipping damage. After damaged components were repaired, the analyzer was calibrated and put in service. Sample probe pluggages, which rendered the SO_2 analyzer inoperable, became a recurring problem. Two system revisions were made to improve probe reliability. The pressure of the blow-back air to the probes was increased to improve probe cleaning and shields were installed around the probes. These revisions did minimize plugging but after a period of continuous operation the shields deteriorated. This problem was not resolved during the first operating season before summer shutdown. Operation of the analyzer system has been good during the 1975-1976 heating season after revisions were made to the probe assembly and purge cycle by the instrument supplier.

Water Balance

The process is designed for closed loop operation, and therefore should have zero scrubbing solution discharge during steady state operating conditions. Extended periods of zero discharge were not achieved during the system start-up period for the following reasons: (1) Excessive water was added to the system thru process pump seals and vacuum filter belt wash sprays. (2) The boilers were often operated at low loads, which resulted in low water evaporation rates in the scrubber.

The loss of scrubbing solution was alleviated by installating rotometers on process pump seal water supply lines and by using clarified scrubber feed solution for vacuum filter belt wash. The system requirement for fresh water was significantly reduced by the installation of a seal water recirculation system on the vacuum pumps. Closed-loop operation has been satisfactorily demonstrated since these modifications have been made.

Freeze Protection

The system as originally installed, did not provide adequate protection from freezing temperatures. Frozen process liquor lines and control air lines contributed to varying degrees of abnormal operation. Installation of additional insulation and higher quality heat tracing has provided reliable freeze protection.

Feed Water Pump

Feed water pump impeller damage due to cavitation was experienced as a result of insufficient suction head in the surge tank which serves as a reservoir for the scrubber feed water pumps. The surge tank height was increased to provide additional system surge capacity and to provide greater suction head to the scrubber feed water pumps. This modification has been very effective in preventing cavitation.

Vacuum Filter Belts

The system has two vacuum filters, each capable of handling the total system solids load. The filter belt originally used was woven of multifilament polypropylene. Belt life was poor due to the accumulation of solids in the interstices of the cloth. It was thought that the belt wash sprays were ineffective, so higher pressure water was piped to the sprays. To date, this modification appears to have prolonged belt life, but other belt materials and cloth weaves are being tried in an effort to extend belt life.

The system was put in service on October 24, 1975, for the 1975-1976 heating season. After approximately two days of operation, the sludge pump impellers began plugging with pieces of compacted sludge that had the consistency of stiff clay. The entire process was idle from June until October, and apparently the solids that could not be removed by filtration during shutdown had compacted on the clarifier walls and rake mechanism. When the process was restarted, this material fell from the metal surfaces, entered the sludge pump, and greatly reduced pump capacity. This situation had never been encountered prior to the first summer shutdown, and a revised shutdown procedure will prevent a problem of this nature in the future.

The lime slaker required considerable maintenance and attention during the first operating season, so hydrated lime was substituted for pebble lime to decrease the maintenance requirement on this equipment. Operation with hydrated lime has been good from a maintenance standpoint, but the hydrated lime has different transport characteristics from the pebble lime, and control of lime feed rate has been poor. Modifications to the volumetric lime feeder have greatly improved lime feed control.

Miscellaneous Hardware Performance

Most of the auxiliary equipment and controls have given good service and have performed adequately. Operating experience and familiarity with the system hardware has improved component reliability through a better understanding of equipment capability and improved preventive maintenance.

Rubber lined diaphragm valves have given excellent service in the process, and have replaced ball valves in some areas.

The radiation type density meters intended to indicate clarifier underflow slurry density were removed due to drift. Slurry density is now monitored by the operator.

Addition of a polyelectrolyte flocculant has yielded the expected favorable results. Clarifier overflow clarity is excellent, slurry filterability has improved, and preliminary results indicate that the cake solids content has been increased from about 54% to around 62% by weight.

The automated sludge conveyor which alternately fills waste containers, has performed well with little operator attention. Periodic cleaning is required to remove sludge from idlers during extremely cold weather.

The Dustraxtor scrubbers, Figure 2, have been reliable and stable even during rapid boiler load fluctuations. The entrainment type scrubber has shown no tendency to scale or plug under any operating conditions. Some soft solids accumulation has been found at the liquid surface in the lower portion of the scrubber, but this does not increase pressure drop or decrease operating efficiency. The solids appear to dissolve away at a rate which prevents any significant accumulation. The demisting section of the scrubbers is bare-metal clean with no solids accumulation whatever.

The difficulties with the lime slaking and lime feed control systems caused some scale formation in the lime mix tanks during the first operating season Improved process control has eliminated this problem.

IV. PROCESS EFFICIENCY AND ECONOMICS

PROCESS EFFICIENCY

A comprehensive system performance test program has just been completed, and only preliminary results of these tests are available. Preliminary performance tests conducted by using the installed SO_2 analyzer and independent wet chemistry analyses indicated that SO_2 removal capability of the process is in excess of 85%. Particulate removal capabilities are yet to be determined.

PROCESS ECONOMICS

A prime reason for selecting the double alkali system was that it afforded an economic advantage over the use of low sulfur coal or conversion to low sulfur oil. Precise cost determinations are difficult due to variable costs of raw materials and associated services, but the best current estimates of system operating costs are shown below:

Basis: 4% S coal
75% SO_2 removal required

	Per Ton of Coal	
Chemicals (lime and sodium carbonate)	$2.56	
Water	0.02	
Electricity	0.71	
Waste	0.64	
Subtotal		$3.93
Labor (operation and maintenance)	$2.12	
Maintenance	0.38	
		$2.50
Depreciation		$3.90
Estimated Total Cost Per Ton Coal		$10.33

V. SUMMARY

The Zurn Double Alkali System at the Joliet plant of Caterpillar Tractor Co. was an advance ment in the technology of flue gas desulfurization and involved certain unknowns. However, the system was made operational on schedule and has encountered few problems, none of which haven't been effectively solved. The process is currently in its second operating season and reliability continues to steadily improve. The system has completed approximately 5600 hours of operation, including 2500 hours of satisfactory continuous operation prior to planned summer shutdown.

System development and optimization are continuing to maximize the potential of the process. In the near future, the remaining minor mechanical problem will be resolved and more data will be available on system performance.

ZURN DOUBLE ALKALI
DESULFURIZATION PROCESS

RAW MATERIALS - UTILITIES USAGE

	PER HOUR	PER TON COAL	PER 1000 LB. STEAM
LIME	1040 LB.	108 LB.	5.78 LB.
SODA ASH	285 LB.	29.5 LB.	1.58 LB.
WATER	36 GAL.	3.7 GAL.	0.2 GAL.
ELECTRICITY	6.85 KWH	0.71 KWH	0.038 KWH
WASTE DISPOSAL	5200 LB.	539 LB.	28.9 LB.

BASIS: 180,000 #/HR. STEAM GENERATED

9.65 T/HR. COAL FIRED

4% SULFUR COAL

75% SO_2 REMOVAL REQUIRED

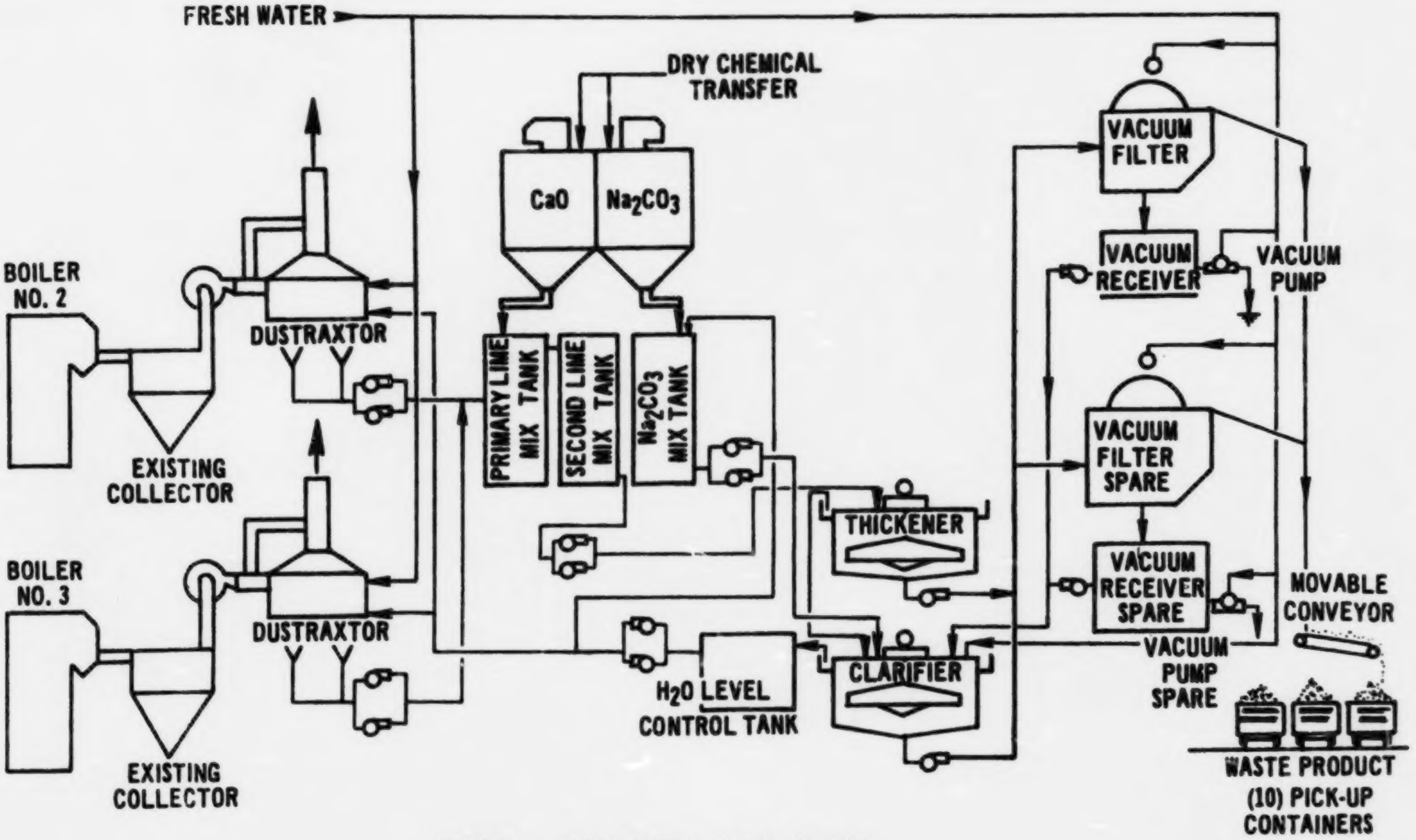

FRESH WATER
DRY CHEMICAL TRANSFER
CaO
Na_2CO_3
BOILER NO. 2
BOILER NO. 3
DUSTRAXTOR
DUSTRAXTOR
EXISTING COLLECTOR
EXISTING COLLECTOR
PRIMARY LIME MIX TANK
SECOND LIME MIX TANK
Na_2CO_3 MIX TANK
THICKENER
CLARIFIER
H_2O LEVEL CONTROL TANK
VACUUM FILTER
VACUUM RECEIVER
VACUUM PUMP
VACUUM FILTER SPARE
VACUUM RECEIVER SPARE
VACUUM PUMP SPARE
MOVABLE CONVEYOR
WASTE PRODUCT (10) PICK-UP CONTAINERS

FIGURE ZURN DOUBLE ALKALI SYSTEM

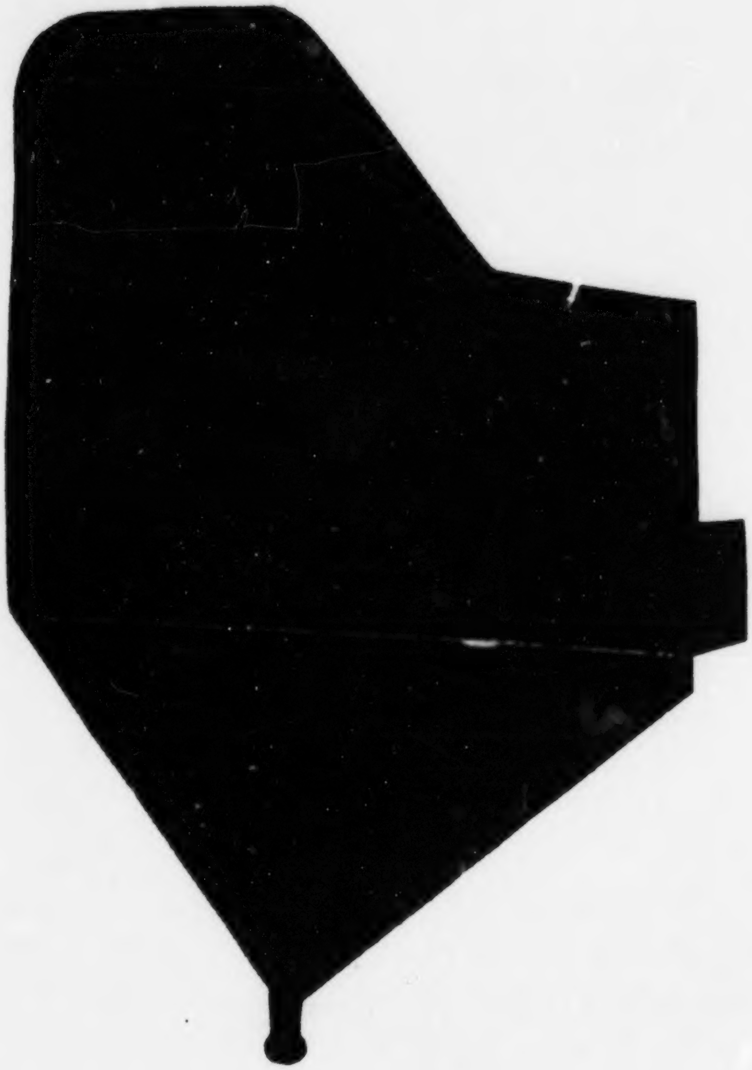


FIGURE 2

ZURN DUSTRAXTOR SCRUBBER

KUREHA FLUE GAS DESULFURIZATION
"SODIUM ACETATE-GYPSUM PROCESS"

Shigeru Saito, Tomijiro Morita, and Shigeyuki Suzuki

Kureha Chemical Industry Co., Ltd.
1-8 Nihonbashi Horidome-Cho, Chuo-Ku,
Tokyo, Japan

ABSTRACT

This paper deals with a new sodium acetate-gypsum desulfurization process. This process is a complete closed loop with a SO_2 removal efficiency of more than 99% using limestone or lime as regenerants and can be applied to flue gases from oil fired as well as coal fired boilers.

Furthermore, a new simultaneous removal of SO_2 and NO_2 is also described.

KUREHA FLUE GAS DESULFURIZATION

" SODIUM ACETATE-GYPSUM PROCESS "

1. INTRODUCTION

Several years ago, Kureha first developed a sodium sulfite process as a single alkali process to produce sodium sulfite for the pulp and paper industries. Then, based on this development, Kureha has further developed a sodium sulfite-gypsum double alkali process in cooperation with Kawasaki Heavy Industries, Ltd. The first full scale plant of this process has been in operation at Shin Sendai power station of Tohoku Electric Power Co., Inc. since April 1974.

Recently, it has also been proven by two full scale plants of Shikoku Electric Power Co., Inc. having around 740,000 st.cu.ft./min. of flow rate.

Through this experience, we have learned that this double alkali process is superior to other existing processes with respect to SO_2 removal efficiency, operability and economy.

However, the processes that we have developed, have been considered mainly for flue gases from oil fired power plants using sodium hydroxide or calcium carbonate as feed materials to produce sodium sulfite or gypsum as by-products.

On the other hand, there are strong demands for the removal of SO_2 and particulate in the flue gas from coal fired boilers as well as flue gas from sinter plants of steel mills.

Also, lime can, in some cases, be utilized more advantageously than can lime stone. There is a strong incentive to develop a process to meet the above requirements.

Key differences in the nature of flue gases from coal fired boilers lie in the higher oxygen content, particulates and chlorine when compared to flue gases from oil fired boilers.

For the development of a new process, these factors have been taken into our serious consideration and acetic acid has been adopted as the best absorption liquor to satisfy the basic requirement for a scrubber absorbent:

1. It should give a weak alkaline solution with large buffering capacity, when its aqueous solution is neutralized by strong base.

2. It should be a weaker acid than sulfurous acid, but stronger acid than carbonic acid.

3. It should have high solubility in water, and the vapor pressure should be as low as possible.

4. Solubility of its calcium salt in water should be large.

5. It should be easily available in the market and the price should be reasonable.

Acetic acid thus selected, has been utilized successfully as an efficient absorbent in the process development through stepwise tests from 60 st.cu.ft./min. bench scale test to 3,000 st.cu.ft./min. pilot scale test for the flue gas from oil fired power stations.

Additionally, we have carefully investigated the feasibility for flue gas from a coal fired boiler using 60 st.cu.ft./min. test facility.

Through a series of tests, we could overcome problems that had been encountered in the cleaning of flue gases from coal fired boilers. We have also obtained the necessary information for the scale-up of this process.

We are confident that this process is superior in operability and economy to the other known processes. Also, in the present situation in Japan, hydrated lime can be used advantageously as a regenerant for this process over lime stone from overall process economy.

We are now in the planning stage of several full scale plants in Japan to demonstrate the superiority of this sodium acetate gypsum process using either limestone or lime for flue gases from oil fired as well as coal fired boilers.

We wish to report the results of our process development of the sodium acetate-gypsum process. 1) 2)

2. PROCESS DESCRIPTION

2-1. Lime Stone Method

As shown in Figure 1, this process is composed of three basic sections; absorption, oxidation and gypsum formation. This process has been tested at 3,000 st.cu.ft./min. pilot test facility for the flue gas from our oil fired power station from March 1975 through August 1975 with continuous and successful operation.

2-1-1. Absorption Process As illustrated in Figure 1, the absorption tower consists of two sections - SO_2 absorption and acetic acid elimination section - , which are linked in series in one absorption tower. The lower section is composed of two absorption chambers and the higher acetic acid elimination section consists of three chambers, each being equipped with perforated plates inside of the chambers.

In the SO_2 absorption section, SO_2 is removed by circulating a sodium acetate solution through the two stages.
The following reaction takes place:

$$2CH_3COONa + SO_2 + H_2O \rightarrow Na_2SO_3 + 2CH_3COOH \qquad (1)$$

A part of the acetic acid thus formed volatilizes in the scrubbed flue gas, and must be eliminated in the next step.

In this step, fresh lime stone slurry is added to the top chamber, and flows down countercurrently from chamber to chamber to the bottom to accomplish the complete removal of acetic acid vapor. Also, this step serves to complete the elimination of remaining SO_2 in the flue gas from the absorption section, so that clean exhaust gas is released into the atmosphere from the top of the scrubber.

For the smooth operation of the scrubber, prevention of scaling is essential. Thus, the following four measures have been taken against scaling prevention.

1. The structure of the scrubber has been designed in such a way as there can be formed no dry parts inside the scrubber walls.

2. Perforated plates have been used successfully in preventing scaling, because vigorous mixing of flue gas and absorption liquor through the plates have resulted in both preventing scaling to and removing scaling from the walls or the plates.

3. Seed crystals have been added successfully to prevent scaling by the possible precipitation of calcium sulfate in the absorbent. Especially, this precipitation tends to take place,when the solubility of calcium sulfate in the absorption liquor becomes lower at the outlet of the absorption section than at the inlet.

4. When the concentration of sodium sulfite is relatively high in the liquor, this reacts with calcium sulfate in the liquor to form insoluble calcium sulfite, which may cause a scaling problem. Accordingly, it is necessary to keep the concentration of sodium sulfite in the absorption liquor lower than 2% as can be understood from Figure 2. This can be successfully accomplished in

our process by diluting the scrubber absorbent with a part of oxidized liquor from the oxidizer.

By the application of these measures, we have achieved the prevention of scaling in the scrubber and at the same time a high efficiency of SO_2 removal and acetic acid elimination for a variable loadings of gas flow rate as summarized in Table 1.

2-1-2. Oxidation Process Sodium sulfite formed in the SO_2 absorber is oxidized to sodium sulfate in the oxidation tower which is provided with perforated plates to facilitate fine dispersion of air bubbles and to promote oxidation. Sulfite oxidation to sulfate takes place as follows:

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4 \qquad (3)$$

After oxidation, the liquor is sent to the gypsum formation section, where gypsum is produced by the addition of lime stone slurry.

2-1-3. Gypsum Production Process In this step, gypsum is produced from the oxidized liquor containing mainly sodium sulfate and acetic acid by the addition of lime stone slurry. In this gypsum formation, calcium carbonate reacts very rapidly with sodium sulfate in the presence of acetic acid, whereas the reaction of calcium carbonate with sodium sulfate is very slow in the absence of acetic acid.

In this process, calcium carbonate is assumed to react at first with acetic acid present in the liquor to form

calcium acetate, which reacts further with sodium sulfate to form calcium sulfate and sodium acetate by a double decomposition reaction. The regenerated sodium acetate is recirculated to the scrubber after separation from gypsum by filtration.

The reaction can be expressed as in (4) and (5):

$$CaCO_3 + 2CH_3COOH \rightarrow (CH_3COO)_2Ca + H_2O + CO_2 \qquad (4)$$

$$(CH_3COO)_2Ca + Na_2SO_4 \rightarrow CaSO_4 + 2CH_3COONa \qquad (5)$$

Gypsum produced in this step has such chemical composition as represented in Table 2.

This gypsum is high in quality and is utilized in wall board making and in Portland cement production.

Thus, use of acetic acid as an absorbent has made it possible to remove SO_2 efficiently from flue gas, producing high quality of gypsum as a by-product.

Also, rapid reaction of sodium sulfate with calcium carbonate has made this process further possible to apply it to flue gases with high oxygen content such as flue gases from coal fired boilers, which will be mentioned in the latter section in detail.

2-2. Acetate-Lime Process

In order to make a comparison between lime stone and lime, tests have been conducted at 60 st.cu.ft./min. bench test facility to compare the date with those obtained by lime stone process.

In principle, acetate-lime (hydrated lime) process has the same structure with the acetate-limestone process as given in Figure 3. The absorbent containing sodium acetate is in a similar way transferred from the absorption tower through the oxidation process to the gypsum formation step, where lime is added instead of lime stone to produce gypsum. This reaction takes place more rapidly than with lime stone even in a very dilute lime suspension. Also, the mother liquor obtained after separation from gypsum crystals shows a higher P^H value than in the case of lime stone process. These two features make it possible to operate the acetate-lime process with advantage over lime stone process especially in terms of removal efficiency of SO_2 and acetic acid removal in the scrubber.

At the same time, concentration of the absorption liquor can be reduced considerably to minimize the possible loss of acetic acid by the oxidation.

Through these experiments, we have found that the scrubber unit for SO_2 can be simplified from two absorption sections to single section and as to acetic acid vapor removal from three to one section, respectively. Also, make-up of acetic acid has been reduced considerably.

In the simplified scrubber, the elimination section of acetic acid by aqueous lime solution played only a supplementary role, because most of acetic acid vapor in the flue gas had been eliminated in the first SO_2 absorption section by the absorption liquor of high P^H value (7.0 - 7.2). Figure 4 supports the above fact, in which relationships between P^H values of the liquor and concentrations of acetic acid in the gas phase are shown. For a P^H range of 7.0 to 7.2 with a temperature of 131°F, concentration of acetic acid in the vapor phase can be reduced as low as 3 parts per million. Encouraged by the above test results, we are planning to prove the operability and economy of this lime process by 3,000 st.cu.ft./min. pilot plant in the near future.

3. DESULFURIZATION OF FLUE GAS FROM COAL FIRED BOILER

Technogical requirements for the desulfurization of flue gases from coal fired boilers are increasing from the standpoint of environmental protection especially in other countries than in Japan such as in the United States or in Germany, where they have big coal reserves as an energy source for power generation. Also, an increase in power generation from imported coals can be also anticipated in the future in Japan.

In respect to these considerations, we have made a train of desulfurization tests for flue gas from coal fired boiler at the 60 st.cu.ft./min. test facility using sodium acetate-lime stone process in order to solve problems, which may encounter in the desulfurization of flue gases from coal fired boilers. These tests were carried out in August of 1975, and the results are reported in the next section.

3-1. Sodium Acetate-Lime Stone Process

Following test items have been considered for the evaluation of the process.

1. Efficiency of desulfurization.

2. Efficiency of particulate removal.

3. Quality of gypsum.

4. Construction material.

5. Material and heat balance.

6. Behavior of trace materials and their accumulations in the system.

7. Confirmation of the system for a closed loop without disposal water.

A flue gas containing 600 parts per million of SO_2 and $1.4x10^{-3}$ to $2.8x10^{-3}$ gr./st.cu.ft. of particulate was introduced into scrubber with an inlet gas temperature of above 100°C.

As to the efficiency of SO_2 removal, nearly complete removal of SO_2 has been achieved as listed in Table 3.

Accordingly, as far as efficiency of desulfurization is concerned, the experimental results are excellent.

As to the removal of particulate in the flue gas, tests have been made for variable contents of particulate and different numbers of perforated plates in the scrubber. The results are shown in Table 4.

In these tests, flue gas containing a high level of particulate (more than 0.14 gr./st.cu.ft.) has been prepared by adding fly ash to the flue gas to determine the removal efficiency. (s. Exp. No. 9 and 10 in Table 4.)

As seen in Table 4, efficiency of particulate removal is remarkable even to the high contents of particulate and a distinct relationship can be seen between particulate removal efficiency and the number of perforated plates used. As to particulate elimination, better results can be anticipated for the scrubber with 12 perforated plates than with an electrostatic precipitator.

During the above tests, no noticeable troubles have been experienced with particulate or chlorides even in the case of high particulate loadings.

The quality of gypsum formed was satisfactory for further processing, even if it had colored gray to light brown depending on the operational conditions.

No influence has been observed in the crystal growth of gypsum, size of which ranged from 100 to 150 microns in the long axis and 50 to 60 microns in the short axis.

As to materials of construction for the test, only corrosion-resistant metals and plastic materials were used, so that no abnormality has been observed during the tests.

Regarding the material and utility balance of this process, a comparison was made between designed values and experimentally obtained values using the 60 st.cu.ft./min.

bench scale test facility. Comparative data are listed in Table 5.

As can be seen from Table 5, smaller experimental values than designed have been obtained as the material balance for lime stone and gypsum, which can probably be ascribed to smaller SO_2 content in the flue gas than designed at the early stage of the test. Otherwise, an excellent coincidence between experimental and projected values have been attained both for material and utility balance.

Regarding impurities such as particulate and chlorine, tests have been made extensively. Acid insolubles in the gypsum were analyzed and found to be composed mainly of silica and alumina, which amount to about 4% of the gypsum. This suggests that the particulate in the flue gas is removed from the system together with gypsum. Thus, in our tests no build-up of particulate has been observed during operation.

As to chlorine content in the flue gas, measurements showed chlorine content of 0.7×10^{-3} to 1.0×10^{-3}gr./st.cu.ft. No operational troubles have been caused by this level of chlorine content, and also, no influence has been observed upon the absorption of SO_2 and acetic acid in the scrubber, the rate of oxidation and of gypsum formation. However, if a closed system is adopted, accumulation of chlorine will take place in the circulating liquor, and must be eliminated from the circulating system. Extensive research is now being made by us to solve this problem.

As far as this bench scale test was concerned, a complete closed loop has been successfully demonstrated without waste water disposal. In this bench scale test, make up of acetic acid and sodium hydroxide were not necessary.

Besides the above items mentioned, acetic acid vapor in the flue gas from the SO_2 scrubbing section has been measured both by gas chromatography and by gas detection tube. The former method gave an acetic acid concentration of 1.9 to 3.5 parts per million in the outlet gas of the scrubber, while the concentration was less than 1 part per million for the same samples by the latter method, showing practically complete removal of acetic acid vapor in the flue gas.

As to pressure drop, a total pressure drop of 11 in. H_2O has been recorded for this process, which corresponds to a mean drop of 2.17 in. H_2O for each chamber of the scrubber. (The scrubber consists of 2 chambers of SO_2 absorption and 3 chambers of acetic acid elimination).

For each perforated plate, a mean pressure drop amounts to 0.51 in. H_2O. This magnitude of pressure drop for perforated plate can guarantee good distribution of flue gas in the large scale scrubber. This was confirmed by another separate experiment by using test facility of 18,000 st.cu.ft./min. gas flow rate. Smooth flow down of the liquor and good dispersion of flue gas into the counter-current liquor through the perforated plates were observed in the above test.

To determine the possible trouble in the acetic acid eliminator by the mist of the absorbent from the absorption section, careful observation has been made, but no troubles have been found.

Scaling in the scrubber was prevented effectively by the measures previously noted, and therefore no scaling problem has been raised during the test.

Through a series of tests made on the flue gas from coal fired boiler, we have succeeded in collecting necessary information for the scale-up of this process, and it will be soon demonstrated to verify the process by 3,000 st.cu.ft./min. pilot plant, followed by demonstration of full scale plants.

4. ECONOMY OF THE PROCESS

On the basis of our experience and on a series of bench and pilot tests, process economics can be estimated with good accuracy.

In Table 6, a standard desulfurization cost of sodium acetate-lime process for the flue gas rate of 588,500 st.cu.ft./min. is listed for 1976 costs in Japan. This cost should be considered as a measure of our sodium acetate - lime process, because it is strongly dependent on the impurity content of raw materials and on the dealing of gypsum, whether it is disposed or utilized. In case gypsum is thrown away, process can be simplified and both the investment cost and utilities cost can be reduced, resulting in the reduction of desulfurization cost.

It should be further noted, that desulfurization cost is greatly influenced by local conditions. However, as can be deduced from Table 6, this sodium acetate-lime process is low in desulfurization cost and excellent in operability.

5. SIMULTANEOUS REMOVAL OF SO_2 AND NO_x BY SODIUM ACETATE-GYPSUM PROCESS.

Taking advantage of sodium acetate-gypsum process, we have further developed a process to eliminate SO_2 and NO_x simultaneously. NO_x removal is furnished in the above system by the addition of a catalyst, which is commonly available in the market for a reasonable price.

In this process, NO_x is reduced in the presence of the catalyst to sodium imidodisulfonate by the sodium sulfite in the absorbption liquor, and sodium imidodisulfonate formed is further decomposed by calcium nitrite and lime in the presence of sulfuric acid to nitrogen and gypsum.
As shown in Figure 5, a part of mother liquor containing sodium imidodisulfonate, is drawn as a side stream from the recirculation line to the scrubber and then, it is decomposed by the above chemicals to nitrogen and gypsum in the reaction vessel.

Accordingly, this system can eliminate SO_2 and NO_x simultaneously by additional attachment of NO_x removal section to the sodium acetate-gypsum process in a complete closed loop without no disposal water.

This process has been proven by 3000 st.cu.ft./min. pilot test and based on this successful results, we are planning to build a full scale plant in the near future.

6. CONCLUSION

Based on long experience of research and development in the field of flue gas desulfurization, we have finally come to sodium acetate-lime process for the treatment of flue gases from both coal and oil fired boilers, process economics can be assumed to be the most advantageous one. Furthermore, we have recently developed a simultaneous removal of SO_2 and NO_x process on the basis of the sodium acetate-gypsum process.

Though these new systems are still on the way of pilot test, we are sure that these systems will be demonstrated in a full scale plant in the near future as one of the most favorable process for various type of flue gases containing SO_2 and particulate.

References

1) T. Kaminaga and K. Noguchi

" Kureha Sodium Acetate-Gypsum FGD Process and Plant "
Chemical Factory (Japan) 19, 5, 85-88 (1975)

2) S. Saito

" Kureha Sodium Acetate-Gypsum FGD Process "
Chemical Engineering (Japan) 20, 3, 17-19 (1975)

Table 1, SO_2 REMOVAL EFFICIENCY BY SODIUM ACETATE-GYPSUM PROCESS AT 3000 st.cu.ft./min. PILOT TEST FACILITY.

Exp.No.	Gas Flow Rate (st.cu.ft./min.)	SO_2 Concentration		SO_2 Removal Efficiency (%)
		Inlet(parts/million)	Outlet(parts/million)	
1	2940	1500 - 1700	less than 1	more than 99.9
2	2350	1500 - 1700	less than 1	more than 99.9
3	1470	1500 - 1700	less than 1	more than 99.9
4	940	1500 - 1700	less than 1	more than 99.9

Table 2, LIME STONE AND GYPSUM QUALITY

Lime Stone

Particle Size	
200 mesh pass	95%
Chemical Composition	
SiO_2	0.3%
Al_2O_3	0.05%
Fe_2O_3	0.07%
CaO	54.3%
MgO (as Mg)	0.3%

Gypsum

Free Water	5.59% (Wet base)
Crystal Water	20.75% (Dry base)
CaO	30.41%
SO_3	45.75%
Purity	96.88% (as $CaSO_4 \cdot 2H_2O$)
Drystal Size	50 μ x 100 μ

Table 3, REMOVAL EFFICIENCY OF SO_2 BY SODIUM ACETATE-LIMESTONE PROCESS IN 60 st.cu.ft./min. BENCH SCALE TEST FOR FLUE GAS FROM COAL FIRED BOILER.

Exp. No.	Concentration of SO_2 (parts/million)		Desulfurization
	Gas Inlet	Gas Outlet	(%)
1	646	0	100.0
2	690	1	99.8
3	570	0	100.0
4	550	1	99.8
5	570	0	100.0
6	580	2	99.6

Table 4, RELATIONSHIP BETWEEN NUMBERS OF PERFORATED PLATES AND PARTICULATE REMOVAL EFFICIENCY FOR A FLUE GAS WITH A FLOW RATE OF 60 st.cu.ft./min.

Exp.No.	No. of Plates Used	Amount of Particulate (gr./st.cu.ft. x 10^3) Inlet	Outlet	Efficiency of Removal (%)
1	3	1.50	0.90	40
2	3	2.66	1.02	62
3	3	2.95	1.16	61
4	5	3.31	0.88	74
5	5	3.28	0.85	74
6	12	3.93	0.48	88
7	12	5.83	0.51	91
8	12	4.50	0.23	95
9	12	2.71	0.28	90
10	12	209.1	2.04	99
11	12	205.6	1.47	99

Table 5, MATERIAL AND UTILITY BALANCE OF SODIUM ACETATE-LIMESTONE PROCESS

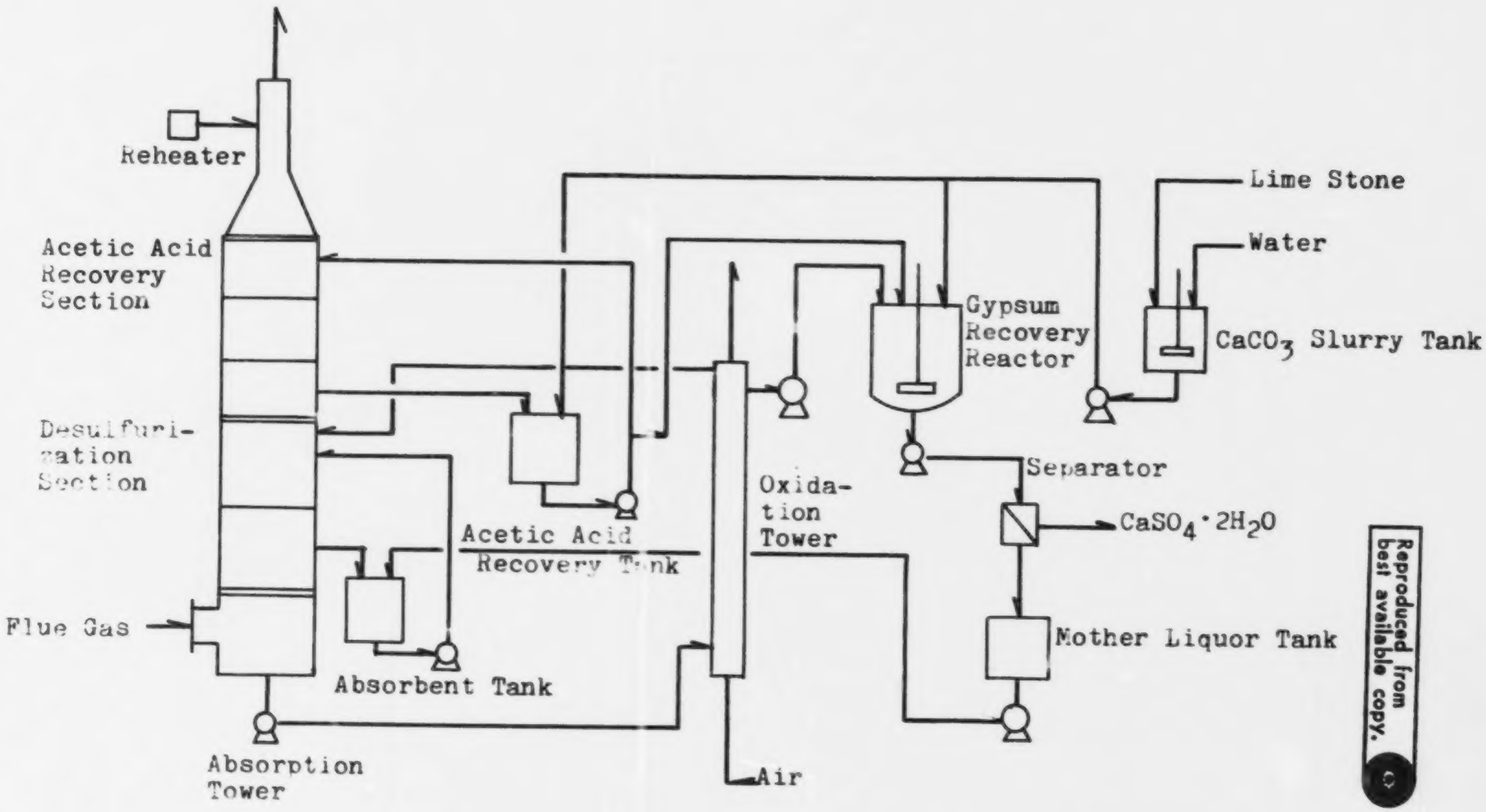
		Experimental	Designed
Gas Composition			
SO_2	Inlet	580-630 parts/million	600 parts/million
SO_2	Outlet	0 - 2 "	1 "
Acetic Acid	Outlet	1.6-3.5 "	3 "
Raw Material			
$CaCO_3$		0.53 lb/hr.	0.59 lb/hr.
By product			
Gypsum		0.88 lb/hr.	1.01 lb/hr.
Utilities			
Process water		0.42 gal./hr.	0.42 gal./hr.
Steam		44.1 lb/hr.	176.3(max.)-22.0(min.)lb/hr.
Power		10 kWh	40 kWh
Air		141.2 st.cu.ft/hr.	141.2 st.cu.ft./hr.

Table 6, ECONOMIC EVALUATION OF SODIUM ACETATE-LIME PROCESS

Flue Gas Volume:	588,500 st.cu.ft./min.
SO_2 Content of Flue Gas:	1,000 parts/million
Efficiency of Desulfurization:	more than 99%
Operating Hours:	8,400 hrs./year.
Investment Cost (BL):	15,800 ¥/KW.
Cost of Desulfurization:	12.57 ¥/gal.
Raw Materials and Utilities	
Hydrated Lime:	7,429.5 lb/hr.
NaOH :	35.3 lb/hr
Acetic Acid:	201.6 lb/hr.
Steam:	8,766.2 lb/hr.
Process Water:	13,210.0 gal/hr.
Electricity:	4,500 KW
Gypsum:	172,840 lb/hr.

Man Power

2 Men/Shift, 4 Shifts/day + 2 Men.


Reheater
Acetic Acid Recovery Section
Desulfurization Section
Flue Gas
Absorption Tower
Absorbent Tank
Acetic Acid Recovery Tank
Oxidation Tower
Air
Gypsum Recovery Reactor
Separator
$CaSO_4 \cdot 2H_2O$
Mother Liquor Tank
Lime Stone
Water
$CaCO_3$ Slurry Tank

Reproduced from best available copy.

FIGURE 1, SCHEMATIC FLOW DIAGRAM OF LIME STONE DESULFURIZATION PROCESS

FIGURE 2, SOLUBILITY OF CALCIUM SULFATE IN THE ABSORPTION LIQUOR IN DEPENDENCE ON SODIUM SULFITE CONCENTRATIONS

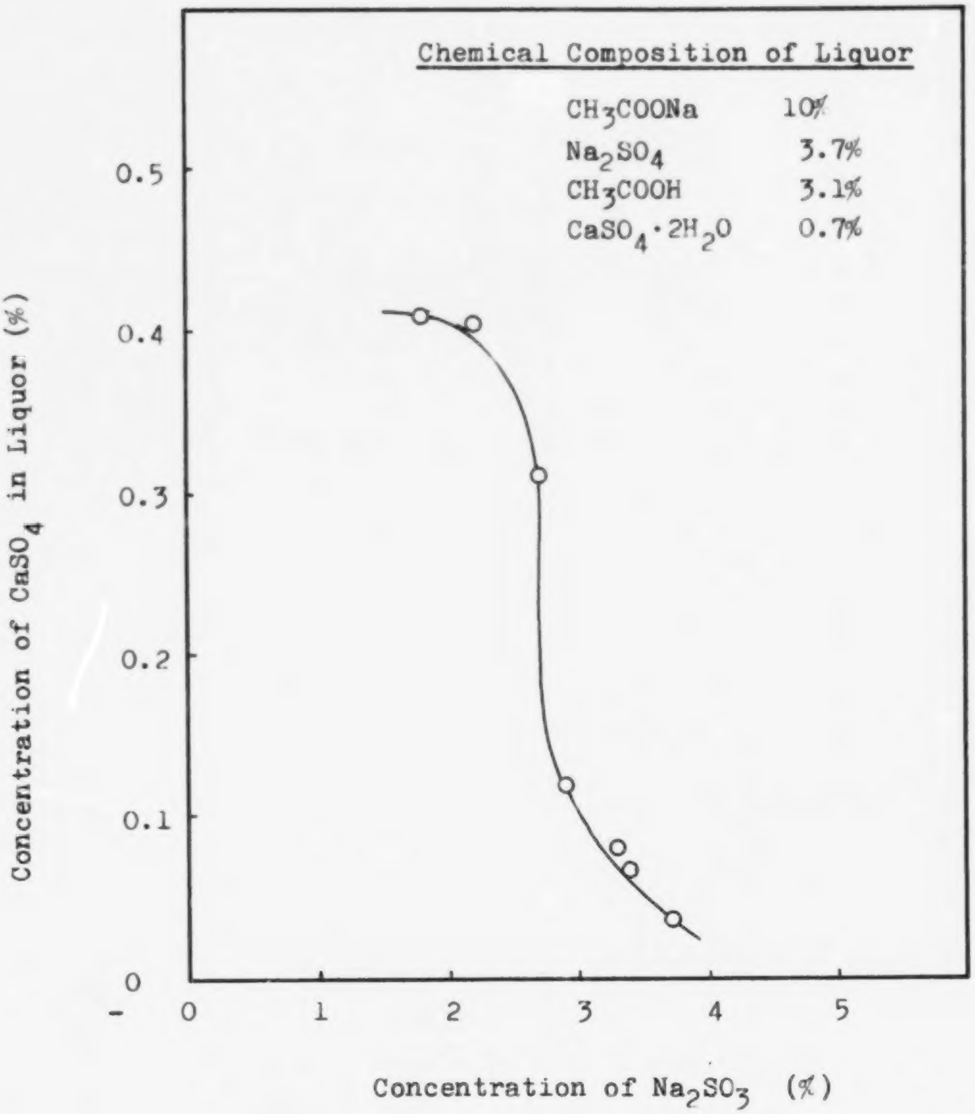


FIGURE 3, LIME DESULFURIZATION PROCESS

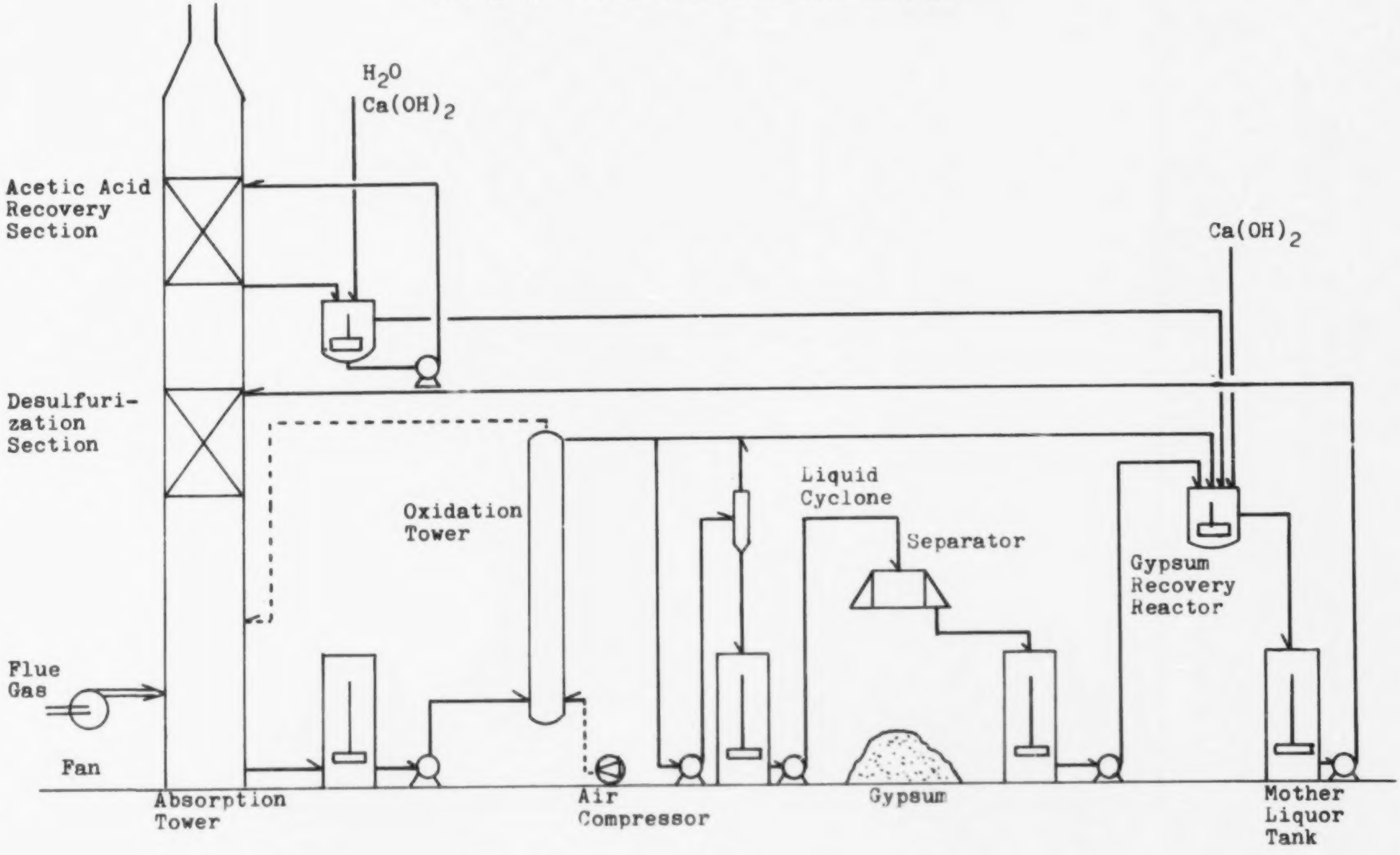

H_2O
$Ca(OH)_2$
Acetic Acid Recovery Section
Ca(OH)$_2$
Desulfuri-zation Section
Oxidation Tower
Liquid Cyclone
Separator
Gypsum Recovery Reactor
Flue Gas
Fan
Absorption Tower
Air Compressor
Gypsum
Mother Liquor Tank

FIGURE 4, RELATIONSHIP BETWEEN P^H OF THE LIQUOR AND ACETIC ACID CONCENTRATION IN THE VAPOR PHASE.

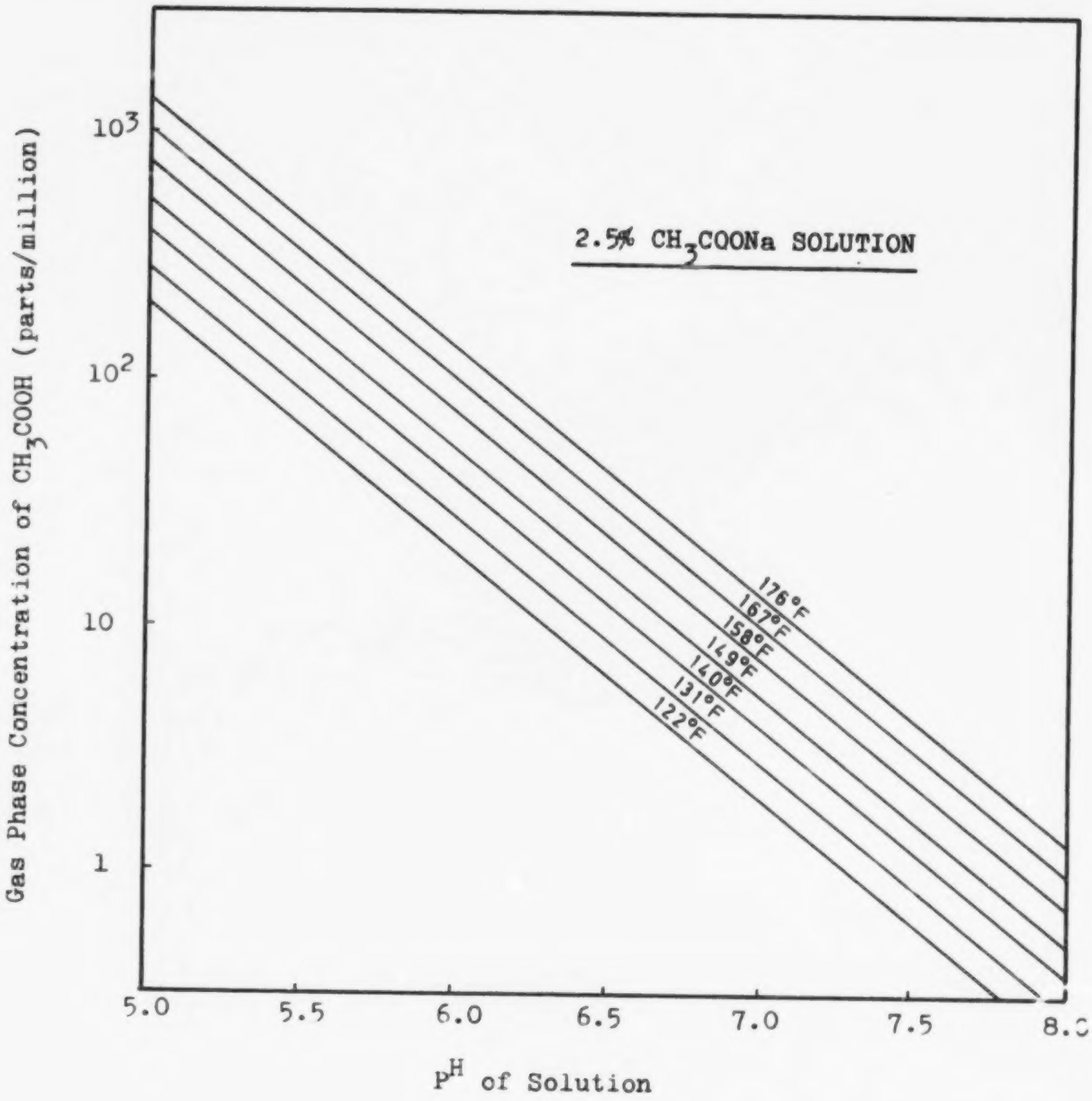
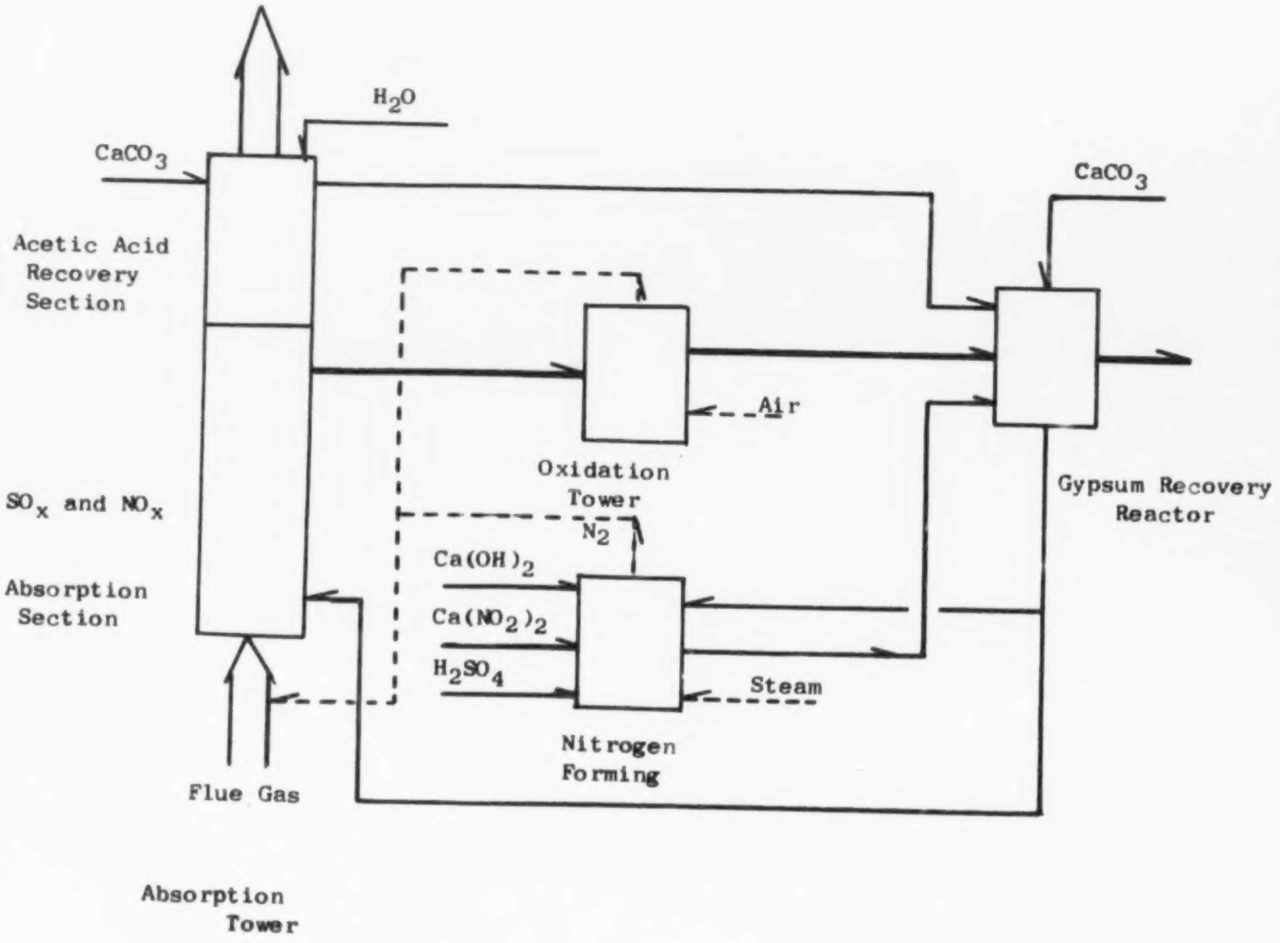


FIGURE 5. SIMULTANEOUS REMOVAL OF SO_2 AND NO_x BY SODIUM ACETATE-GYPSUM PROCESS.


H_2O
$CaCO_3$
$CaCO_3$
Acetic Acid Recovery Section
Air
Oxidation Tower
Gypsum Recovery Reactor
SO_x and NO_x
N_2
$Ca(OH)_2$
Absorption Section
$Ca(NO_2)_2$
H_2SO_4
Steam
Nitrogen Forming
Flue Gas
Absorption Tower

THE BUELL DOUBLE-ALKALI SO_2 CONTROL PROCESS

H. Edward Bloss

Buell-Envirotech
Lebanon, Pennsylvania

James Wilhelm

EIMCO-Envirotech
Salt Lake City, Utah

William J. Holhut

Central Illinois Public Service Company
Springfield, Illinois

ABSTRACT

This paper traces the development of the Buell Double-Alkali SO_2 Control Process to the current technology of concentrated and dilute mode operation that are currently offered commercially. Areas of application are defined for the two modes, each resulting in elimination of chemical scaling, abrasive scrubbing slurries and high maintenance expenditure. Design information is also presented for utilities service with high chloride coals. The Central Illinois Public Service Company, Newton Station, Unit #1 (concentrated mode, high chloride pulverized coal operation) currently under design is presented and equipment defined. Preliminary capital investment and operating cost estimates are tabulated.

Preceding page blank

INTRODUCTION

Scrubbing of sulfur dioxide from boiler flue gas has been a major problem for the utilities industry due to costs, complexities, and impact on reliability of electric generating operations. An important solution to this problem lies in continued advancement of process design technology which will achieve required reduction in stack SO_2 emission without reducing generating unit availability and without creating secondary environmental pollution effects that may jeopardize the quality of surface and ground water bodies.

Because of variations in boiler installations and fuel sources, a single scrubbing system design has not been able to achieve the objectives of all applications. For that reason, Envirotech has invested more than four years work in developing and evaluating the applicability of new process design technology to control the various operating variables encountered with operating boiler units.

ENVIROTECH'S APPROACH

As a means of minimizing capital investment, the utilities industry continues to focus on flue gas desulfurization systems of the waste "throwaway" type. In view of their comparatively low cost and their ability to precipitate collected sulfur oxides as a water-insoluble solid, calcium-containing raw materials such as lime or limestone are solely used to convert the collected SO2 waste to the solid form. The basic double-alkali system shown on Figure 1 accomplishes the above in two advantageous steps, scrubbing first, post-precipitation later:

Equation 1 (Absorption)

$SO_2 + Na_2SO_3 + H_2O \qquad 2\ NaHSO_3$

Equation (1a.) $SO_2 + H_2O \qquad H_2SO_3$

Equation (1b.) $H_2SO_3 + Na_2SO_3 \qquad 2\ NaHSO_3$

and/or

Equation 2 (Absorption)

$SO_2 + H_2O + 2\ NaOH \qquad Na_2SO_3 + 2H_2O$

Equation 3 (Regeneration)

$$2\ NaHSO_3 + Ca(OH)_2 \qquad Na_2SO_3 + CaSO_3.1/2\ H_2O + 3/2\ H_2O$$

and/or

Equation 4 (Regeneration of fully oxidized collected sulfur oxides)

$$Na_2SO_4 + Ca(OH)_2 + 2\ H_2O \qquad CaSO_4.2\ H_2O + 2\ NaOH$$

All sodium-base double-alkali systems perform generally within these equations.

Beginning with initial work five years ago, Buell and Eimco have committed efforts to the advancement of the double-alkali design approach to SO_2 control technology. The throwaway SO_2 system double-alkali has a special advantage in avoiding precipitation of the SO_2 solids in the gas-handling scrubbing system and instead, applies the lime treatment externally of a clear liquor scrubbing circuit that utilizes highly water-soluble sodium chemicals as shown in equations above.

Envirotech's work in double-alkali flue gas desulfurization has developed two similar flow sheets that are now being offered or applied in commercial applications. The first, the dilute mode process - is specifically designed for boilers operating with abnormally high excess air or for low sulfur coal applications such as in western U.S.A. The second, the concentrated mode process - is used for high sulfur coal applications in pulverized coal fired boilers where the excess air in the flue gas is generally low enough to limit the oxidation of the absorbed sulfur species, as required for its zero-effluent operation.

DILUTE MODE PROCESS

For power plants that have high excess air and/or low flue-gas SO_2 concentrations, the oxidation rate of the absorbed SO_2 changes the chemical equilibrium relationships in the system. The major alkali in the process is sodium hydroxide not sodium sulfite and the calcium ion concentration in the regenerated liquor is as high as 500 ppm. In order to control chemical scaling of the scrubber due to intrusion of calcium via the regenerated liquor return stream, a calcium softening step is utilized as shown in Figure 2. Soda ash makeup required by the system is added to this softener, or Reactor-Clarifier, to precipitate soluble calcium and eliminate scaling potential in the scrubber. The liquor flow volumes in the post-precipitation step of this system are dictated by the concentration of alkali (NaOH) that can be formed in the regenerated liquor under conditions of chemical equilibrium governed by Equation 4, above.

The normal ionic concentrations for this system are 0.1 molar NaOH and 1 molar Na. This variation of the basic double-alkali system provides scale control and process reliability but has a captial cost higher than a comparable concentrated mode (if it were applicable), due to the equipment for control of the soluble calcium and the higher volume of liquor that must be handled by the post-precipitation system.

CONCENTRATED MODE PROCESS

In the concentrated mode (concentrated active-alkali) process, the flue gas is scrubbed with a concentrated sodium sulfite/bisulfite solution to remove sulfur oxides. A bleed stream of spent scrubbing liquor at a pH near 6.0 is reacted with lime to precipitate calcium sulfite and the limited amount of calcium sulfate that has been formed due to oxidation in the scrubber. The precipitated solids are then thickened, filtered and water-washed (for minimization of sodium losses), to produce a disposable waste for landfill disposal, and the regenerated liquor is returned to the scrubber to absorb additional SO_2. The normal ionic concentrations for this system are approximately 0.5 molar Na_2SO_3 and 2 molar Na. This flow sheet is shown in Figure 3.

Due to limitations in the calcium sulfate precipitation kinetics, this system is used where the oxidation of the SO_2 may be kept below about 25% of the total absorbed SO_x species. The magnitude of this oxidation rate depends upon the ratio of the SO_2 absorbed to the excess air or oxygen in the flue gas. In general, this ratio must be at least 20 ppm of SO_2 per excess air percentage in order to adequately control the oxidation rate. As a typical example, a boiler operating at 30% excess air requires a minimum of 600 ppm or about 1% sulfur in the coal to insure that oxidation levels are sufficiently limited. At sulfur levels greater than the 600 ppm in gas, or 1% in fuel, effective precipitation of the sulfate ions is assured.

This double-alkali system removes at least 90 to 95% to the SO_2 in the flue gas. A relatively dry filter cake waste is the sole waste product. There are no liquid effluents discharged from the process. Scrubber scaling is controlled because the regenerated liquor contains a very low calcium ion concentration (20-80 ppm) because of the limited aqueous solubility-product constant of the calcium sulfite species. The scrubbing solution is a clear liquor (containing only a small amount of residual fly-ash), that has essentially no abrasive characteristics and cannot produce internal scaling.

HIGH CHLORIDE PROCESS

Absorption of hydrogen chloride from the flue gas can cause special problems in all throwaway systems including the double-alkali process. As indicated above, filter cake washing is employed to wash out and recycle the soluble sodium salts to the process. This procedure minimizes the amount of soda ash required for make-up. When chloride is absorbed into the scrubbing liquor, sodium chloride concentration-buildup in the system is magnified by this washing and recovery of the soluble salts from the filter cake. In some cases, augmental sodium chloride concentrations as high as 1.5 to 2.0 molar may result, causing sodium sulfate insolubility problems as well as increased usage of make-up soda ash, with higher sodium content in the filter cake.

In order to eliminate these problems, a modification to the basic flow sheet has been developed to remove the chlorides in a separate precooler section upstream from the SO2 absorber. This flow sheet is shown in Figure 4. The chlorides and other secondary coal contaminants are removed in a low pH scrubbing liquor. A slipstream of the resulting acidic solution is neutralized with lime or limestone to produce a calcium chloride solution which is used as a final wash on a horizontal filter of the Eimco-Extractor type. This wash is so designed that the filter cake absorbs the chlorides.

The net results of this chloride removal loop when applied to high-chloride coals (approximately 0.1% chloride by weight in coal or higher), are to:

1. Reduce sodium carbonate make-up requirements; thus reduce operating costs to normal double-alkali system levels.

2. Reduce the dissolved solids concentration in the SO_2 scrubbing system; thus to reduce salting out problems and achieve normal optimum double-alkali process operation.

3. Reduce the amount of sodium salts in the filter cake; thus producing more suitable landfill waste material.

4. Eliminate most of the residual fly-ash and other secondary coal contaminants from the SO_2 removal system. This minimizes the amounts of heavy metal ions in the SO_2 scrubbing liquor and thus reduces the catalytic effects of these ions in promoting oxidation of the sulfite and bisulfite ions.

In a further refinement of this process, a fly ash thickener is added to the precooler loop. Thus, the system can be used as a back-up to the electrostatic precipitators and still operate at a low suspended solids level with low abrasion characteristics in the precooler liquor. The primary function of this additional thickener is to accommodate temporary malfunctions of the electrostatic precipitators such as shorts of electrical sets and temporary power loss. In the interim, while the back-up function is actually being fulfilled, the precooler system is not subjected to abrasion since solids build-up and related problems have been avoided: the system will continue to function as designed. This is the basic flow sheet that has been accepted and is being applied for use at the Newton Station of Central Illinois Public Service Company, scheduled for 1977 startup of operations. The system, currently in the engineering phase, is the concentrated mode/high chloride process with auxiliary fly-ash thickener loop to control the system during precipitator upsets.

PLANT DESCRIPTION & DESIGN CONDITIONS

The double-alkali system will be installed as an adjunct to the Newton Station, Unit #1, Newton, Illinois. The boiler is a drum-type manufactured by Combustion Engineering, Inc., designed for a maximum steam generating capacity of 4,158,619 lbs/hr. at 2,620 psig and 1,005°F. at the steam outlet connection. The boiler unit and the double-alkali control system are scheduled for in-service operation December, 1977.

The boiler is tangentially fired with pulverized coal. The pulverizing mills are the bowl-type manufactured by Combustion Engineering, Inc. Net theoretical air required for combustion at maximum steaming condition is 4,192,000 lbs/hr at 60°F. and 80% RH. The boiler is designed for operation at 22% excess air. High sulfur Illinois bituminous coals from several sources may be burned. In the design of the system, the following flue gas analysis was developed from the worst case criteria:

	LBS/HR.
SO_2	37,800
H_2O	330,000
O_2	556,000
N_2	4,580,000
CO_2	1,110,000
SO_3	700
HCl	1,000
TOTAL	6,615,000 (1)

The system is designed for coal properties and boiler system characteristics as follows:

Sulfur Content	4.0% in coal
Chloride Content	0.2% in coal
Coal Heating Value as received	10,900 BTUS/lb.
Boiler Heat Release	5.5×10^9 BTUS/hr.

95% of Sulfur in fuel Reports as SO_2 in flue gas
96% of Cl in fuel Reports as HCl in the flue gas

(1) Value represents total I.D. fan capacity including excess and in-bleed air leakage.

PROJECT SCOPE

Buell is responsible for the design, procurement and delivery to the job site, of a double-alkali flue gas desulfurization facility (FGDU) designed for 90% removal of SO_2. The station design includes two induced fans which discharge into a common plenum.

Each fan is also connected to the stack. The double-alkali flue gas desulfurization unit, which is being built to accommodate Unit 1, will include the following major equipment and facilities:

A. Four booster fans.

B. The ductwork, including the plenum, leading to the four booster fans, also including isolation dampers for each fan.

C. Isolation gate dampers in the boiler ductwork to isolate the I.D. Fans from the stack.

D. Four precooler towers.

E. Ductwork from the booster fans to the four precoolers.

F. Four scrubber towers.

G. Mist eliminators between the precoolers and the scrubbers.

H. Mist eliminators at the outlet of the four scrubber towers.

I. Ductwork from each of the four scrubbers connecting them to two plenums, each of which is common to two scrubbers. This ductwork will be complete with outlet isolation gates for each scrubber.

J. All ductwork to connect the scrubber system back to the chimney.

K. Lime unloading and handling system including:

    1. Equipment designed to receive pebble or granular quick-lime using both rail and truck unloading facilities.

2. Four lime storage silos for the storage of pebble or granular lime.

3. Lime feeding and ball-mill-slaking facilities for the addition of lime to the reaction tanks.

4. A building sheltering the lime feeding and slaking facilities.

L. Soda ash system, consisting of:

1. Soda ash off-loading facilities designed to receive soda ash from both rail and truck.

2. Two soda ash storage silos.

3. Soda ash feeding and dissolving equipment.

M. Two 100' diameter thickeners for the double-alkali post-precipitation system with concrete bottoms and access to bottom discharge cone using tunnel.

N. One 50' diameter thickener for the precooler liquor loop, constructed of coated steel and elevated above grade.

O. Three Eimco Extractor filters for dewatering and washing of double-alkali-system thickener underflow, and associated equipment including vacuum pumps, filtrate receivers, filtrate pumps, and moisture traps. A building to house the filters and all related equipment is also included with the FGDU.

P. A control and maintenance building.

Q. Flocculant addition systems for the thickeners.

R. Material off-loading facilities including a train shed (about 250' long) which will accommodate four cars and material handling equipment to unload two cars at one time from a total of six hoppers. The material off-loading facility is able to handle trucks which may or may not have their unloading auxiliaries.

The Project Scope also includes necessary accessary equipment and facilities required to fulfill all performance guarantees.

GENERAL COMMENTS

Design of flange-to-flange wet-end equipment such as reactors, thickeners, vacuum filters, is based on the process bleed liquor flow rates in accordance with established sizing criteria per Envirotech pilot plant experience. Wherever feasible, the Envirotech wet-end system design incorporates gravity flow for operating simplicity and to reduce oxidation, covered troughs have been utilized in place of pipe for gravity runs wherever possible. Materials of construction are generally mild steel lined with appropriate corrosion and/or abrasion resistant materials. Piping is sized for flow rates of 4-7 feet per second. Process pumps are designed per system flow sheet with an added 20% safety factor. Rubber lined pumps with rubber-covered impellers are utilized for corrosive and/or abrasive service. All process pumps are spared where necessary to insure continuity of system operation.

Instrumentation is designed to permit automatic operation of the system, and includes compensation for load changes, and an emergency water deluge system for high-temperature protection of scrubber internals including non-metallic linings. Facilities provide for the recording of key system operating data, i.e., liquid flow rates, temperatures, pressures, pH, inlet and outlet SO_2 concentrations, etc. Manual override is also provided for. In the event of complete system shutdown, provisions have been made to drain all lines and pumps to drain sumps. The drain sumps are then pumped to vessel freeboard space in the post-precipitation system. Tankage is sized to allow for the minimum liquid retention time consistent with optimum process operation taking into account the additional volume needed for liquid storage during system shutdown. Materials of construction are generally non-metallic lined, 1/4" thick, mild steel. A vacuum filtration system is provided and is sized based on the thickener bottoms solids-flow rate. Provision for a spare filter plus variable speed drive permits extra capacity. Should it become necessary to remove the filters from operation for a short period of time, the thickeners are capable of handling an amount of solids build-up. The thickener mechanisms are designed to permit an accumulation (12" - 24") of settled solids in the thickener tanks through the incorporation of automatically operated lifting devices that raise and lower the rakes.

The wet-end (post-precipitation) system incorporates a filter building which houses the vacuum filter system. A separate building houses the process control room. The latter is a dust free, heated and air conditioned room housing the system instrumentation and operators' comfort facilities. The reaction tanks are provided with a launder system that permits by-passing of any one unit for maintenance without shutting down the process.

CAPITAL INVESTMENT AND ANNUAL OPERATING COST ESTIMATES

The project is currently in the design engineering phase and finalized costs have yet to be developed. It is currently estimated that the direct investment will be $23 million with the completed cost of $40 million which would reflect the total capital investment estimated at this phase of the project. Since the investment costs have not been developed at this time for all the individual equipment items, an analysis of the fixed investment cannot be made. However, based on the estimates and the known feedstock/utility and manpower requirements, the following annual operating cost values have been developed:

ESTIMATED AVERAGE ANNUAL OPERATING COSTS

DOUBLE-ALKALI PROCESS

(575 MW NEW COAL FIRED BOILER UNIT, 4.0% IN FUEL; 90% REMOVAL)

DIRECT COSTS	ANNUAL QUANTITIES	UNIT COST $	TOTAL ANNUAL COST $	PERCENT OF TOTAL ANNUAL OPER. COST
Lime (93% Avail. CaO)	114,975 M tons	35/ton	4,024,125	30.9
Soda Ash (98% Na_2CO_3)	3.833 M tons	80/ton	306,640	2.3
Subtotal Raw Materials			4,330,765	33.2

DIRECT COSTS	ANNUAL QUANTITIES	UNIT COST $	TOTAL ANNUAL COST $	PERCENT OF TOTAL ANNUAL OPER. COST
Conversion Costs				
Operating Labor and Supervision	42,924 Man Hours	8.00 per Man Hrs.	343,392	2.6
Utilities				
Steam	53,300 M Lbs.	1.20/M Lbs.	183,960	1.4
Process Water	312,732 M Gal.	0.30/M Gal.	93,820	0.7
Electricity	79,716,000 KWH	0.012 per KWH	956,592	7.3
Maintenance				
Labor and Material (.02 x 40,000,000)			800,000	6.1
Subtotal Conversion Costs			2,377,764	18.3
Subtotal Direct Costs			6,708,529	51.5

INDIRECT COSTS	TOTAL ANNUAL COST	PERCENT OF TOTAL AMOUNT OPER. COSTS
Average Capital Charges @ 14.0% of Total Capital Investment	5,600,000	43.0
Overhead		
Plant	683,553	5.3
Administrative, 10% of Operating Labor	34,339	0.2
Subtotal Indirect Costs	6,317,892	48.5
Total Annual Operating Cost	13,026,421	100.0

	DOLLARS/TON COAL BURNED	MILLS/KWH	CENTS/MILLION BTU HEAT INPUT	DOLLARS/TON SULFUR REMOVED
Equivalent Unit Operating Cost	8.42	3.69	38.62	233.89

ESTIMATE BASIS:

Preliminary Project Cost Estimate

Life of Boiler Plant: 30 years

Coal Burned, 1,547,064 tons/year, 10,900 BTU/hr.

Stack Gas Temperature -154°F. (Combination Steam Reheat & Bypass Mixing)

Boiler Unit on Stream Time, 6,132 hrs. @ Full Load - (70% Load Factor)

Total Capital Investment, $40,000,000 (est'd.); Subtotal direct investment $23,000,000 (est'd.)

CONCLUSION

In order to achieve a flue gas desulfurization system design that results in a high degree of reliability for a specific boiler application, recognition of all key criteria and variables is required. The double-alkali system operating in the concentrated mode is a system design with no scaling potential and minimized abrasion when applied to a modern pulverized-coal utility boiler service. With low-quality coals, with high chloride content, the addition of the separate precooler absorption loop for chloride as well as fly-ash control is a further development for that increased reliability and reduced operating cost.

These systems currently have redundancy and/or excess capacity in their design to accommodate operating unknowns. As technology develops, the redundancy can be reduced and the complexity capacity reduced. With this development, investments cost and operating costs will be reduced and reliability will continue to be achieved.

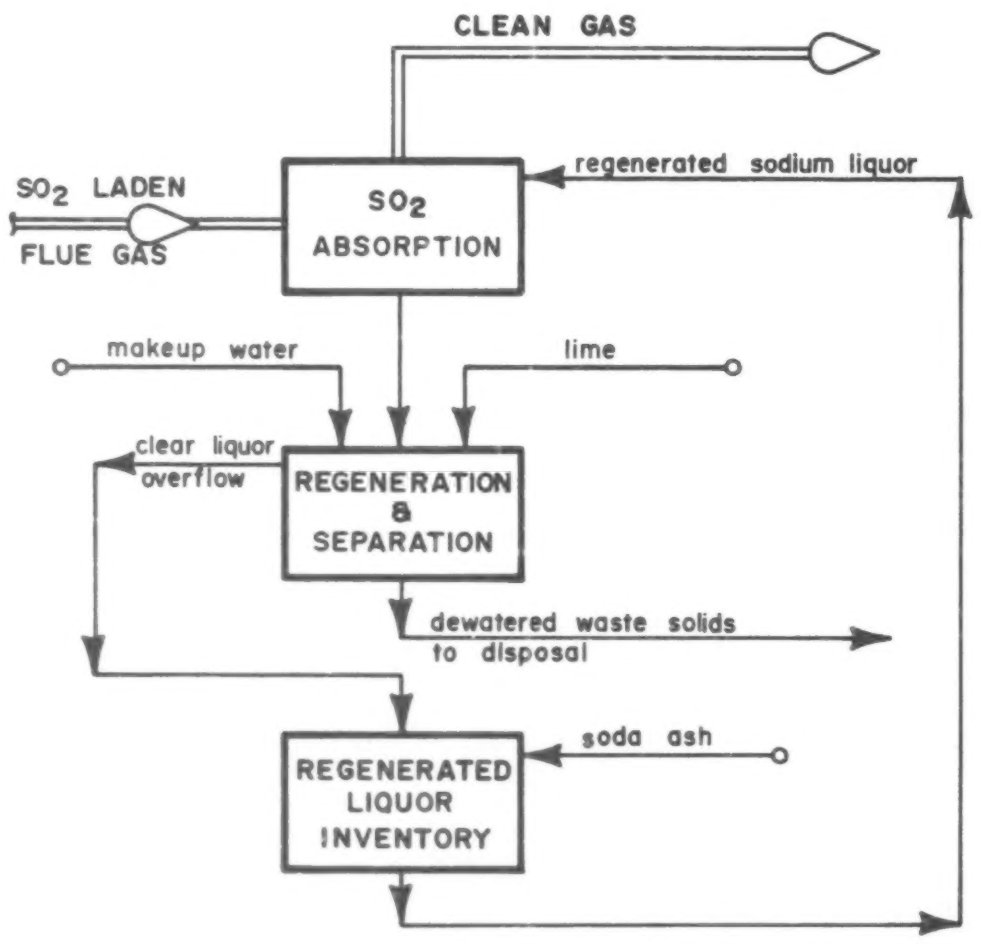

CLEAN GAS
SO_2 LADEN FLUE GAS
SO_2 ABSORPTION
regenerated sodium liquor
makeup water
lime
clear liquor overflow
REGENERATION & SEPARATION
dewatered waste solids to disposal
soda ash
REGENERATED LIQUOR INVENTORY

fig. 1 - BASIC SCHEME

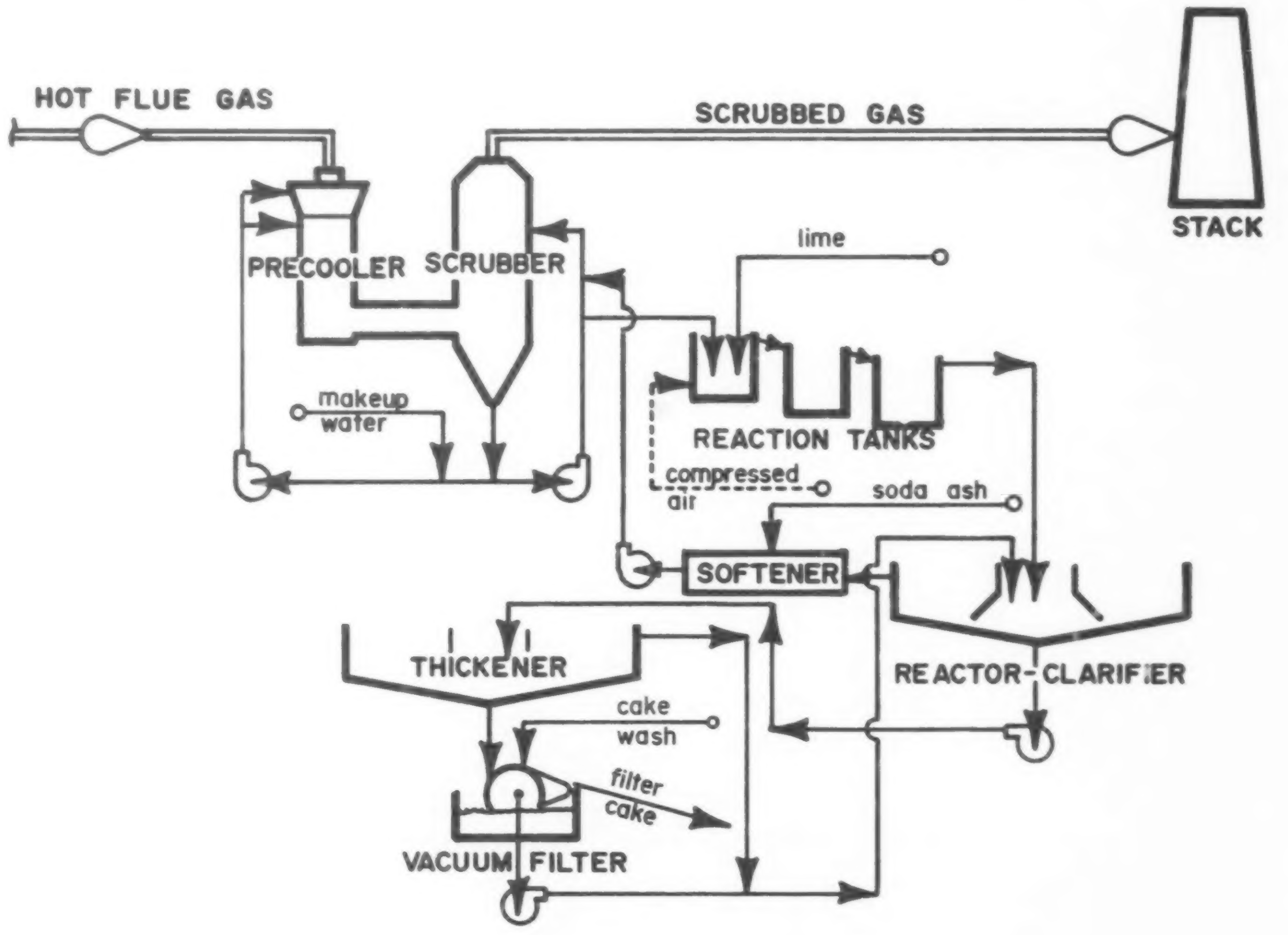

HOT FLUE GAS
SCRUBBED GAS
STACK
PRECOOLER
SCRUBBER
lime
makeup water
REACTION TANKS
compressed air
soda ash
SOFTENER
THICKENER
REACTOR-CLARIFIER
cake wash
filter cake
VACUUM FILTER

fig. 2 - DILUTE MODE

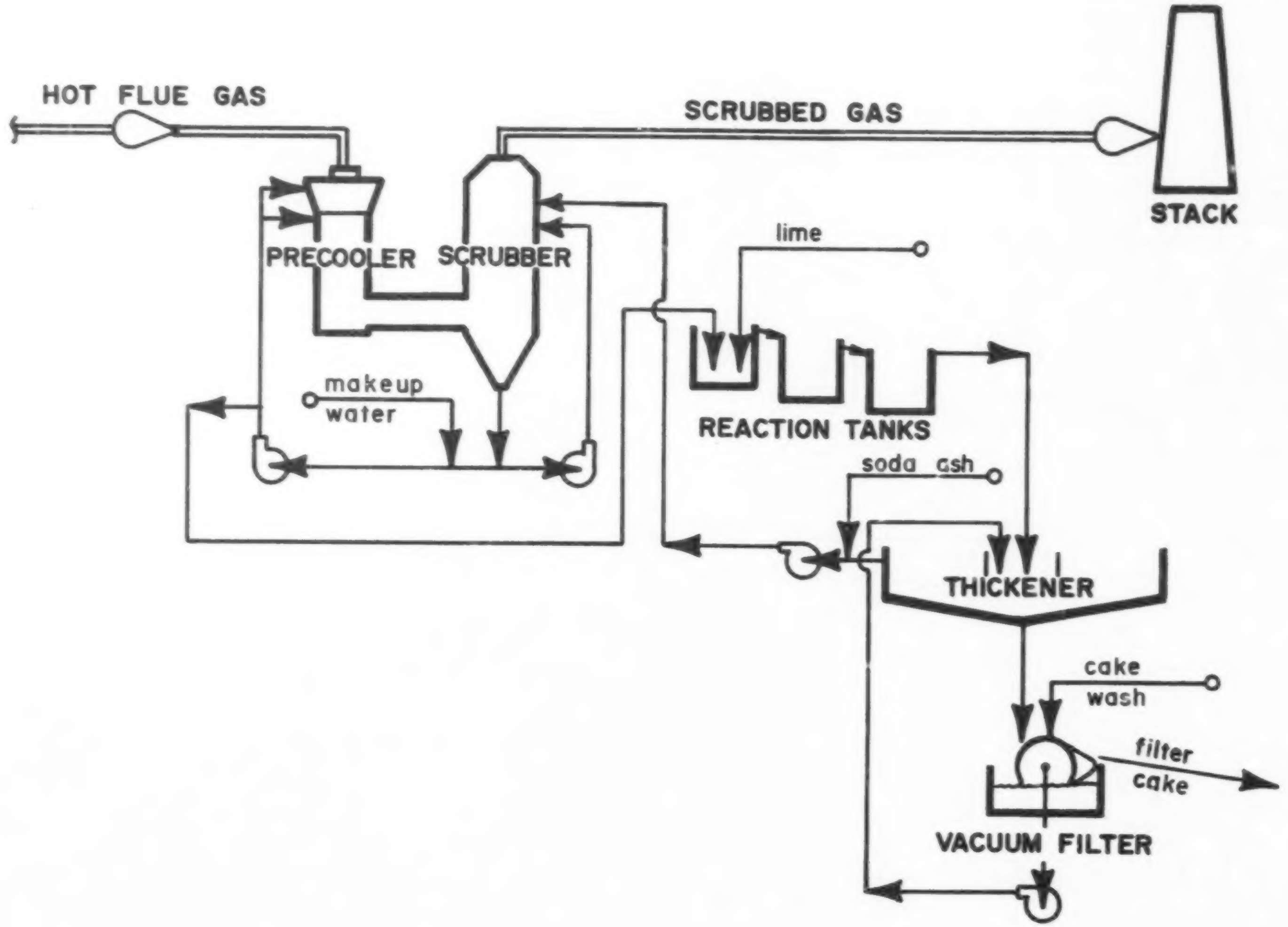

HOT FLUE GAS
SCRUBBED GAS
STACK
PRECOOLER
SCRUBBER
lime
makeup
water
REACTION TANKS
soda ash
THICKENER
cake
wash
filter
cake
VACUUM FILTER

fig. 3- CONCENTRATED MODE

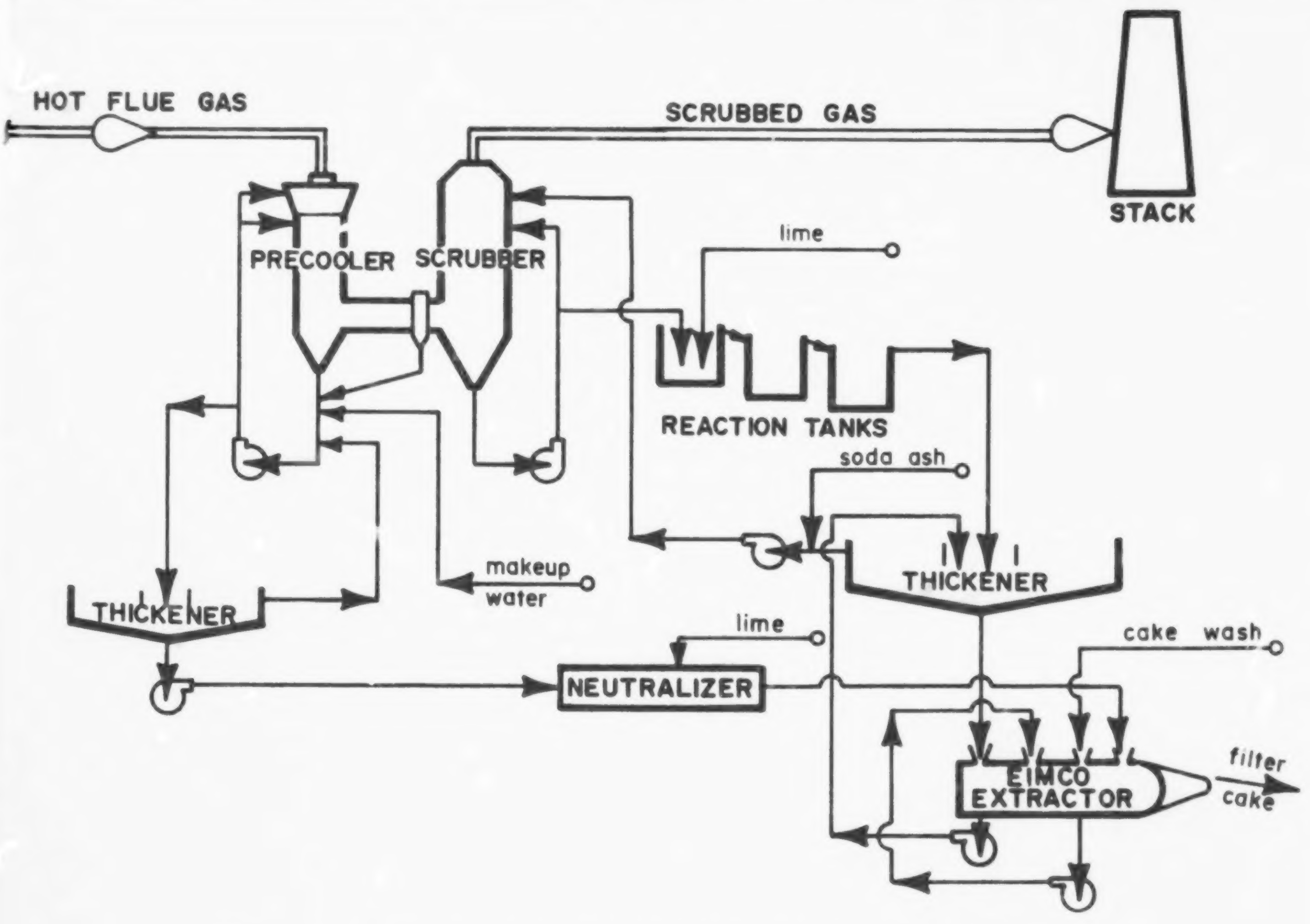

HOT FLUE GAS
SCRUBBED GAS
STACK
PRECOOLER
SCRUBBER
lime
REACTION TANKS
soda ash
THICKENER
makeup
water
THICKENER
lime
NEUTRALIZER
cake wash
EIMCO
EXTRACTOR
filter
cake

fig. 4- HIGH CHLORIDE MODE